

POWER SYSTEMS DEVELOPMENT FACILITY
TOPICAL REPORT

GASIFICATION TEST CAMPAIGN TC16

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ABSTRACT

In support of technology development to utilize coal for efficient, affordable, and environmentally clean power generation, the Power Systems Development Facility (PSDF) located in Wilsonville, Alabama, routinely demonstrates gasification technologies using various types of coals. The PSDF is an engineering scale demonstration of key features of advanced coal-fired power systems, including a KBR (formerly Kellogg Brown & Root) Transport Gasifier, a hot gas particulate control device, advanced syngas cleanup systems, and high-pressure solids handling systems.

This report discusses Test Campaign TC16 of the PSDF gasification process. TC16 began on July 14, 2004, lasting until August 24, 2004, for a total of 835 hours of gasification operation. The test campaign consisted of operation using Powder River Basin (PRB) subbituminous coal and high sodium lignite from the North Dakota Freedom mine. The highest gasifier operating temperature mostly varied from 1,760 to 1,850°F with PRB and 1,500 to 1,600°F with lignite. Typically, during PRB operations, the gasifier exit pressure was maintained between 215 and 225 psig using air as the gasification oxidant and between 145 and 190 psig while using oxygen as the oxidant. With lignite, the gasifier operated only in air-blown mode, and the gasifier outlet pressure ranged from 150 to 160 psig.

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TABLE OF CONTENTS

<u>Section</u>	<u>Page</u>
Inside Cover	
Disclaimer	
Abstract	
Acknowledgement	
List of Tables	iii
List of Figures	iv
1.0 EXECUTIVE SUMMARY	1-1
1.1 Summary.....	1-1
1.2 Test Campaign Objectives.....	1-1
1.3 Test Campaign Summary.....	1-1
1.4 Test Campaign Performance.....	1-2
1.4.1 Transport Gasifier Performance.....	1-2
1.4.2 Particulate Control Device Performance.....	1-3
1.4.3 Performance of Other Systems.....	1-4
1.5 Conclusions.....	1-4
2.0 OPERATIONS	2.1-1
2.1 Project Description.....	2.1-1
2.2 Detailed Test Campaign Objectives.....	2.2-1
2.3 Detailed Test Campaign Summary.....	2.3-1
2.4 Inspections	2.4-1
2.4.1 Transport Gasifier.....	2.4-1
2.4.2 Particulate Control Device	2.4-1
3.0 TEST CAMPAIGN PERFORMANCE	3.1-1
3.1 Transport Gasifier Performance.....	3.1-1
3.1.1 Overview	3.1-1
3.1.2 Gas Composition	3.1-1
3.1.3 Syngas Heating Values	3.1-3
3.1.4 Gasifier Solids Analyses	3.1-5
3.1.5 Carbon Conversion.....	3.1-11
3.1.6 Gasification Efficiencies	3.1-13

3.2 Particulate Control Device Performance.....	3.2-1
3.2.1 Overview	3.2-1
3.2.2 Particle Mass Concentrations	3.2-2
3.2.3 Real Time Particle Monitoring	3.2-3
3.2.4 PCD Solids Analysis	3.2-3
3.2.4.1 Particle Size Distributions.....	3.2-3
3.2.4.2 Dustcake and Bridging Observations	3.2-4
3.2.4.3 Physical Properties and Chemical Compositions	3.2-5
3.2.4.4 Dustcake Flow Resistance	3.2-6
3.2.5 Baseline Pressure Drop Analysis	3.2-7
3.2.6 Backpulse Optimization Testing.....	3.2-8
3.2.7 Filter Element Mechanical Testing.....	3.2-8
3.3 Performance of Other Systems.....	3.3-1
3.3.1 Piloted Syngas Burner/Combustion Turbine.....	3.3-1
3.3.2 Fuel Cell.....	3.3-1
3.3.3 Steam/Oxygen Eductor.....	3.3-1
3.3.4 Coal Feed Systems	3.3-1
3.3.5 Advanced Syngas Cleanup.....	3.3-2

APPENDICES

A-1 Operation History	A1-1
A-2 Equipment List	A2-1
A-3 Mass and Energy Balances	A3-1
A-4 Operating Trends	A4-1
A-5 LHV Projection Calculations.....	A5-1

LIST OF TABLES

<u>Table</u>	<u>Page</u>
2.3-1	Typical Operating Conditions for the Transport Gasifier and Particulate Control Device 2.3-10
3.1-1	Operating Periods 3.1-15
3.1-2	Operating Conditions 3.1-17
3.1-3	Raw Gas Composition, Molecular Weight, and Heating Value 3.1-19
3.1-4	Syngas Sulfur Concentration..... 3.1-21
3.1-5	Projected Gas Composition, Molecular Weight, and Heating Value 3.1-23
3.1-6	Coal Analysis..... 3.1-25
3.1-7	Sorbent Analysis 3.1-26
3.1-8	Standpipe Solids Analysis..... 3.1-27
3.1-9	Loop Seal Solids Analysis..... 3.1-28
3.1-10	PCD Solids from FD0540 Analysis 3.1-29
3.1-11	Historical Standpipe and PCD Fines..... 3.1-30
3.1-12	Carbon Conversion and Gasification Efficiencies 3.1-31
3.2-1	PCD Inlet and Outlet Particulate Measurements 3.2-10
3.2-2	Physical Properties of In Situ Samples and Hopper Samples..... 3.2-11
3.2-3	Chemical Composition of In Situ Samples and Composite Hopper Samples..... 3.2-12
3.2-4	Physical Properties of Bridged Deposit and Dustcake Samples 3.2-13
3.2-5	Chemical Composition of Bridged Deposit and Dustcake Samples 3.2-14
3.2-6	Transient Drag Determined from PCD ΔP and from Lab Measurements..... 3.2-15
3.2-7	Comparison of Average Drag Values Determined from PCD Performance and from Lab Measurements 3.2-16
3.2-8	Axial Tensile Results for Pall FEAL Filter Element 27065 3.2-17
3.2-9	Hoop Tensile Results for Pall FEAL Filter Element 27065..... 3.2-17
3.3-1	Fixed Bed Desulfurization Sorbents in Hot Vessels..... 3.3-5
3.3-2	Fixed Bed HCl Sorbent in Hot Vessel 3.3-5
3.3-3	Fixed Bed Ammonia Cracking and Hydrocarbon Reforming Catalyst in Mini Reactor..... 3.3-5
3.3-4	Actual Operating Parameters for Fixed Bed Desulfurization/HCl Removal in Hot Vessels using Syngas 3.3-6
3.3-5	Actual Operating Parameters for Fixed Bed Ammonia Cracking in Mini Reactor using Bottle Gases 3.3-6
3.3-6	Actual Operating Parameters for Ammonia Cracking and Hydrocarbon Reforming in Mini Reactor using Syngas 3.3-7

LIST OF FIGURES

<u>Figure</u>		<u>Page</u>
2.1-1	Flow Diagram of the PSDF Gasification Process	2.1-3
2.1-2	KBR Transport Gasifier	2.1-4
2.1-3	Siemens Particulate Control Device.....	2.1-5
2.4-1	Notch at Entrance to Cyclone.....	2.4-3
2.4-2	Bridged Gasification Ash on Bottom Plenum of PCD	2.4-4
3.1-1	H ₂ O Data.	3.1-33
3.1-2	Wet Syngas Compositions.....	3.1-33
3.1-3	Wet Syngas Molecular Weight and Nitrogen Concentration	3.1-34
3.1-4	Sulfur Concentrations	3.1-34
3.1-5	Sulfur Capture	3.1-35
3.1-6	Syngas Lower Heating Values.....	3.1-35
3.1-7	Raw Lower Heating Value and Overall Percent O ₂ in Feed	3.1-36
3.1-8	Raw Lower Heating Value and Air to Coal Ratio	3.1-36
3.1-9	Raw Lower Heating Value and Oxygen to Coal Ratio.....	3.1-37
3.1-10	Coal Carbon and Moisture	3.1-37
3.1-11	Coal Sulfur and Ash	3.1-38
3.1-12	Coal Heating Value.....	3.1-38
3.1-13	Standpipe Solids SiO ₂ , CaO, and Al ₂ O ₃	3.1-39
3.1-14	Loop Seal Solids SiO ₂ , CaO, and Al ₂ O ₃	3.1-39
3.1-15	Loop Seal Solids Organic Carbon.....	3.1-40
3.1-16	PCD Solids Organic Carbon.....	3.1-40
3.1-17	PCD Solids Silica and Alumina	3.1-41
3.1-18	PCD Solids Calcium Carbonate and Calcium Sulfide	3.1-41
3.1-19	PCD Solids Calcination and Sulfation.....	3.1-42
3.1-20	Standpipe, Loop Seal, and PCD Solids Organic Carbon Content	3.1-42
3.1-21	Gasifier and PCD Solids Calcium	3.1-43
3.1-22	Gasifier and PCD Solids Silica.....	3.1-43
3.1-23	Coal Particle Size	3.1-44
3.1-24	Sorbent Particle Size.....	3.1-44
3.1-25	Percent Coal Fines and Oversize	3.1-45
3.1-26	Standpipe Solids Particle Size	3.1-45
3.1-27	Standpipe Solids Fine and Coarse Particles	3.1-46
3.1-28	Loop Seal Solids Particle Sizes.....	3.1-46
3.1-29	PCD Solids Particle Sizes	3.1-47
3.1-30	Particle Size Distribution.....	3.1-47
3.1-31	Gasifier and PCD Solids Bulk Density.....	3.1-48
3.1-32	Carbon Conversion	3.1-48
3.1-33	Carbon Conversion and Mixing Zone Temperature.....	3.1-49
3.1-34	Carbon Conversion of Four Coals.....	3.1-49
3.1-35	Cold Gasification Efficiency	3.1-50

3.1-36	Hot Gasification Efficiency	3.1-50
3.2-1	PCD Inlet Particle Concentration as a Function of Coal Feed Rate	3.2-18
3.2-2	PCD Outlet Dust Concentration for Recent Gasification Runs	3.2-19
3.2-3	Gasket Removed and Dissected After TC16	3.2-20
3.2-4	Response of PCME Particulate Monitor in TC16 and Earlier Test Runs	3.2-21
3.2-5	Comparison of Large Particle Fraction in TC16 and TC14 PCME Injection Tests.....	3.2-22
3.2-6	Effect of Gasification Oxidant on PCD Inlet Particle Size Distribution with PRB Coal	3.2-23
3.2-7	Effect of Sorbent on PCD Inlet Particle Size Distribution with PRB Coal.....	3.2-24
3.2-8	Effect of Coal Type on PCD Inlet Particle Size Distribution	3.2-25
3.2-9	Particle Size Distribution of PCD Hopper Samples	3.2-26
3.2-10	Specific Surface Area versus Carbon Content of In Situ Samples	3.2-27
3.2-11	Lab Measured Drag as a Function of Particle Size	3.2-28
3.2-12	PCD Transient Drag versus Carbon Content of In Situ Samples.....	3.2-29
3.2-13	Comparison of PCD Transient Drag with Laboratory Measurements	3.2-30
3.2-14	Pressure Rise during Backpulse Optimization Testing	3.2-31
3.2-15	Axial Stress-Strain Curves for Pall FEAL Filter Element 27065.....	3.2-32
3.2-16	Hoop Stress-Strain Curves for Pall FEAL Filter Element 27065	3.2-32
3.2-17	Ultimate Tensile Strength at RT versus Hours in Operation.....	3.2-33
3.2-18	Ultimate Tensile Strength at RT and 750°F versus Hours in Operation.....	3.2-33
3.2-19	Strain-to-Failure at RT and 750°F versus Hours in Operation.....	3.2-34
A1-1	Operating Hours Summary for the PSDF Gasification Process	A1-4
A3-1	Mass Balance	A3-3
A3-2	Energy Balance.....	A3-3
A3-3	Carbon Balance	A3-4
A3-4	Sulfur Balance.....	A3-4
A4-1	Gasifier Mixing Zone, Riser, and Outlet Temperatures, 7/14/04 through 8/4/04	A4-1
A4-2	Standpipe and Loop Seal Temperatures, 7/14/04 through 8/4/04.....	A4-1
A4-3	Gasifier Pressures, 7/14/04 through 8/4/04.....	A4-2
A4-4	Gasifier Differential Pressures, 7/14/04 through 8/4/04.....	A4-2
A4-5	PCD Temperatures, 7/14/04 through 8/4/04.....	A4-3
A4-6	PCD Baseline Pressure Drop and Face Velocity, 7/14/04 through 8/4/04 ...	A4-3
A4-7	System Temperature Profile, 7/14/04 through 8/4/04.....	A4-4
A4-8	System Gas Flows, 7/14/04 through 8/4/04.....	A4-4
A4-9	Main Air Compressor Operation, 7/14/04 through 8/4/04	A4-5
A4-10	Original Coal Feeder Operation, 7/14/04 through 8/4/04	A4-5
A4-11	Sorbent Feeder Operation, 7/14/04 through 8/4/04.....	A4-6
A4-12	Syngas Analyzers, 7/14/04 through 8/4/04.....	A4-6
A4-13	Atmospheric Syngas Combustor Operation, 7/14/04 through 8/4/04.....	A4-7
A4-14	Fluidized Bed Combustor Operation, 7/14/04 through 8/4/04	A4-7

A4-15	Gasifier Mixing Zone, Riser, and Outlet Temperatures, 8/9/04 through 8/26/04	A4-8
A4-16	Standpipe and Loop Seal Temperatures, 8/9/04 through 8/26/04.....	A4-8
A4-17	Gasifier Pressures, 8/9/04 through 8/26/04.....	A4-9
A4-18	Gasifier Differential Pressures, 8/9/04 through 8/26/04	A4-9
A4-19	PCD Temperatures, 8/9/04 through 8/26/04.....	A4-10
A4-20	PCD Baseline Pressure Drop and Face Velocity, 8/9/04 through 8/26/04 .	A4-10
A4-21	System Temperature Profile, 8/9/04 through 8/26/04.....	A4-11
A4-22	System Gas Flows, 8/9/04 through 8/26/04.....	A4-11
A4-23	Main Air Compressor Operation, 8/9/04 through 8/26/04	A4-12
A4-24	Original Coal Feeder Operation, 8/9/04 through 8/26/04	A4-12
A4-25	Developmental Coal Feeder Operation, 8/9/04 through 8/26/04.....	A4-13
A4-26	Sorbent Feeder Operation, 8/9/04 through 8/26/04.....	A4-13
A4-27	Syngas Analyzers, 8/9/04 through 8/26/04.....	A4-14
A4-28	Atmospheric Syngas Combustor Operation, 8/9/04 through 8/26/04.....	A4-14
A4-29	Fluidized Bed Combustor Operation, 8/9/04 through 8/26/04	A4-15

1.0 EXECUTIVE SUMMARY

1.1 SUMMARY

This report discusses test campaign TC16 of the KBR Transport Gasifier train with a Siemens Power Generation particulate control device (PCD) at the Power Systems Development Facility (PSDF) located near Wilsonville, Alabama. The PSDF is a flexible test facility designed to develop advanced coal-fired power system components and assess the associated integration and control issues. TC16 began on July 14, 2004, and lasted until August 24, 2004, accumulating 835 hours of gasification operation.

1.2 TEST CAMPAIGN OBJECTIVES

TC16 evaluated gasifier and PCD operations with Powder River Basin (PRB) subbituminous coal and high sodium lignite from the Freedom mine in North Dakota, using air, enriched air, and pure oxygen as the gasification oxidants. The primary test objectives were:

- Test solid oxide fuel cell with coal derived syngas.
- Operate the gasifier during oxygen blown mode at higher than previously tested lower mixing zone (LMZ) pressure using an eductor.
- Test high sodium lignite at various gasifier conditions.
- Conduct failsafe testing with a new system that utilizes rupture discs to simulate a filter element failure.
- Demonstrate the operational reliability and measurement accuracy of the gas sampling system.

1.3 TEST CAMPAIGN SUMMARY

On July 14, 2004, coal feed began at a low rate to heat the gasifier to 1,650°F. When the gasifier reached normal operating conditions, the new upper mixing zone (UMZ) oxygen nozzles were successfully commissioned with air to evaluate the control valves and instrumentation. On July 15, 2004, PCD backpulse tests were performed and three pressure taps were measured to evaluate the backpulse intensity.

Oxygen-blown operations began on July 16, 2004, and PCD backpulse testing proceeded. Syngas flow to the gas cleanup processes and fuel cell preheat began on July 17, 2004, with syngas flow to the fuel cell beginning on July 18, 2004.

Beginning July 19, 2004, the limestone feeder was operated to maintain the standpipe level. On July 22, 2004, a solids injection test was completed to evaluate the response of the online particulate monitor, the PCME, to further characterized its performance in detecting particle penetration through the PCD. Limestone feed testing for TC16 was also completed on July 22, 2004.

Dolomite addition began on July 24, 2004, in preparation for the lignite run. The hot gas cleanup minireactor heat-up was started, and syngas flow was later initiated.

The piloted syngas burner inlet line was preheated while operating the combustion turbine (CT) in simple cycle on July 29, 2004. The following day, syngas was initiated to the PSB/CT in preparation for turbine stack testing that was performed on August 2, 2004.

On August 3, 2004, coal feed was stopped, and a controlled burn of the PCD filter cake and bridged material was performed. The system was shut down and depressurized to inspect the standpipe screw cooler and the cyclone. A few pieces of refractory and deposited material were found in the screw cooler, and inspections revealed that the notch was reforming at the cyclone inlet.

The startup burner was lit and, after the gasifier was around 1,100°F, coal feed began on August 9, 2004. On August 10, 2004, oxygen-blown testing resumed, the new eductor was successfully commissioned to allow for higher UMZ pressures, and an eductor operating envelope was developed for UMZ pressures up to 200 psig.

On August 17, 2004, the gasifier pressure, temperature, and coal feed rate were reduced and the process was smoothly transitioned to high sodium lignite. Over the next several days, the process operated well and work was performed on automating the level control on the continuous fine ash depressurization (CFAD) system.

After a total of 835 hours of coal feed for TC16, the process was shut down and the system was depressurized at 11:03 p.m. on August 24, 2004. A controlled burn procedure was performed on the PCD filter elements, the remaining systems were stopped, and the gasifier was emptied of solids.

1.4 TEST CAMPAIGN PERFORMANCE

Performance of the major equipment during TC16 is summarized in the following three sections:

1.4.1 Transport Gasifier Performance

- During air-blown gasification using PRB, the raw lower heating values at the exit of the gasifier were between 48 and 65 Btu/SCF, resulting in projected heating values at the turbine inlet of a commercial gasifier of between 117 and 140 Btu/SCF. The projection converts PSDF data into commercial projections by accounting for the use of recycle gas, the lower heat loss per pound of coal fed, and the use of cold gas cleanup in a commercial gasifier. The TC16 results were consistent with the raw lower heating values produced in previous test runs.
- In air-blown gasification using lignite, the raw lower heating values at the exit of the gasifier were between 25 and 40 Btu/SCF, resulting in projected heating values at the turbine inlet of a commercial gasifier of between 103 and 118 Btu/SCF.

- In oxygen-blown gasification using PRB, the raw lower heating values at the exit of the gasifier ranged from 59 to 87 Btu/SCF, resulting in projected heating values between 242 and 263 Btu/SCF.
- The carbon conversion was between 92 and 96 percent for both air- and oxygen-blown gasification using PRB coal and was between 83 and 87 percent using high sodium lignite. The TC16 carbon conversions were typical for PRB coal, but lower than normal for high sodium lignite due to lower operating temperatures.
- The raw cold gasification efficiency ranged from 47 to 55 percent and from 31 to 42 percent during air-blown operations using PRB and lignite, respectively. It ranged from 52 to 61 percent for the oxygen-blown operating periods using PRB. The commercially projected efficiency was between 68 to 72 percent and from 50 to 69 percent during air-blown operations using PRB and lignite, respectively. The commercially projected efficiency was between 74 and 79 percent for the oxygen-blown operating periods using PRB.
- The hot gasification efficiency ranged from 80 and 86 percent and from 67 to 74 percent during air-blown operations using PRB and lignite, respectively. It ranged from 82 to 87 percent for the oxygen-blown operating periods using PRB.
- Sulfur concentrations were between 197 and 348 ppm for air blown with PRB, 292 and 465 ppm for oxygen-blown operation, and 733 and 962 ppm for air blown with lignite.

1.4.2 PCD Performance

- Overall, performance of the PCD was relatively stable throughout both portions of the test run. Bridging occurred during the first portion of the test run (from July 14 to August 3, 2004), but did not recur after restarting. There was one filter element failure on the bottom plenum.
- Outlet loading samples indicated good sealing of the PCD, and the outlet loading was maintained below the detection limit of 0.1 ppmw for most of the test run.
- The rupture disc failsafe tester was successfully tested with a Pall Fuse failsafe. The performance of the failsafe was evaluated and the outlet loading approximately 4 hours after the device opened was less than 0.1 ppmw.
- Backpulse optimization testing was performed during TC16. The backpulse pressures were varied from 150 to 300 psi above system pressure to determine the lower limits of the backpulse settings. The backpulse valve open time was also varied at 0.2 and 0.4 seconds. The differential pressure measurements for the 0.2 second time were slightly higher than the 0.4 second time.
- The physical and chemical characteristics of the gasification ash were generally consistent with those from previous PRB and lignite tests. However, major differences were seen in the gasification ash when the gasifier fuel was switched from PRB coal to high sodium lignite. The bulk porosity was reduced from 90.2 percent with PRB coal to 78.1 percent with lignite. The surface area dropped from 197 m²/g with PRB coal to 97 m²/g with high sodium lignite.

1.4.3 Performance of Other Systems

- A Delphi solid oxide fuel cell operated on syngas generated in the Transport Gasifier for 118 hours, generating a maximum output of 800 Watts.
- To prepare for fuel cell operation, the syngas cleanup units tested desulfurization sorbents, hydrochloric acid sorbents, and a nickel-based catalyst for ammonia and organics cracking. The syngas cleanup units operated for over 570 hours. The gas cleanup units were able to supply the fuel cell with clean syngas that contained less than 1.0 ppm sulfur compounds and less than 1.0 ppm HCl. During the last 72 hours of fuel cell operation, the gas cleanup system was able to reduce the total condensable organic content in the syngas to below 300 ppm by cracking the organics using the nickel-based catalyst at 1,600°F.
- The minireactor also assisted in cracking syngas hydrocarbons. The unit ran for over 500 hours and reduced benzene concentrations from 1,000 to 200 ppm.
- The PSB operated on syngas for around 7 hours, at syngas flow rates up to 13,000 pph. The unit produced 10.3 MWh while on syngas.

1.5 CONCLUSIONS

TC16 was a successful test run accumulating 835 hours of on-coal operation using Powder River Basin (PRB) and Freedom lignite coal. The major highlights included:

- A Delphi solid oxide fuel cell operated on syngas for 118 hours during TC16. The first fuel cell stack ran for 28 hours, during which time the performance declined significantly. Another fuel cell stack was installed and fuel cell performance only degraded slightly during the first eight hours of testing, then remained steady for 82 hours.
- The PSB testing lasted for approximately seven hours at syngas flow rates up to 13,000 pph and the combustion turbine (CT) operated for about 20 hours.
- The steam/oxygen eductor was successfully operated, blending the steam and oxygen to generate a higher supply pressure and allowing oxygen addition at higher gasifier pressures. The gasifier exit pressure was increased from 155 psig up to 200 psig.
- The PCD rupture disc failsafe tester was successfully used with the Pall fuse to simulate catastrophic filter failure. Outlet loading after the test was below the detection limit.
- The continuous fine ash depressurization system (CFAD) worked very well and ran for 834 hours during TC16. The new automatic level control was tested at the end of the test run and also worked well.
- Lignite operations in TC16 were improved over previous testing. Air blown Freedom lignite standpipe solids decreased to 175 microns (SMD) and were steady at 175 microns for the last 80 hours of TC16. This was in marked contrast to the previous testing of high sodium lignite testing in TC13 when the standpipe solids were often higher than 400 microns (SMD).

2.0 OPERATIONS

2.1 PROJECT DESCRIPTION

The Power Systems Development Facility (PSDF), near Wilsonville, Alabama, is funded by the U.S. Department of Energy, Southern Company, and other industrial participants currently including the Electric Power Research Institute, Siemens Power Generation, KBR, Peabody Energy, and the Lignite Energy Council. The PSDF is an engineering scale demonstration of key features of advanced coal-fired power systems designed at sufficient size to evaluate system components and assess the integration and control issues of these advanced power systems. The facility also supports clean coal technology programs to address environmental concerns associated with using fossil fuels for producing electricity, chemicals, and transportation fuels.

The KBR Transport Reactor which operates at the PSDF is a pressurized, advanced circulating fluidized bed reactor which can operate in either combustion or gasification mode. While operating in gasification mode, either air or oxygen can be used as the oxidant. The particulate-laden gas exiting the reactor is filtered by a downstream high temperature, high pressure filter vessel, the particulate control device (PCD). In gasification mode, the objective of the PCD is to clean the gas sufficiently so that it can be utilized in a downstream gas turbine/combustor or fuel cell. A gas clean-up skid is also available to remove various pollutants from a syngas slipstream. A flow diagram of the gasifier train is shown Figure 2.1-1. The Transport Gasifier train has now operated for about 5,000 hours in combustion mode and over 6,100 hours during gasification.

The Transport Gasifier, shown in Figure 2.1-2, consists of a mixing zone, a riser, a disengager, a cyclone, a standpipe, a loopseal, and a J-leg. Steam and either air or oxygen are mixed together and introduced in the lower mixing zone (LMZ) while the fuel, sorbent, and additional air, oxygen, and steam (if needed) are added in the upper mixing zone (UMZ). The steam and oxidant, along with the fuel, sorbent and solids from the standpipe, are mixed together in the UMZ. The UMZ, located below the riser, has a slightly larger diameter than the riser. The gas and solids move up the riser before entering the disengager, which removes larger particles by gravity separation. The majority of the solids flow from the disengager into the standpipe, and the remaining solids flow, along with the syngas, to the cyclone, which removes most of the particles not collected by the disengager. The solids collected by the disengager and cyclone are recycled back to the gasifier mixing zone through the standpipe and a J-leg. The nominal gasifier operating temperature is 1,800°F, and the gasifier system is designed to have a maximum operating pressure of 294 psig with a thermal capacity of about 41 MBtu/hr. Due to a lower oxygen supply pressure, the maximum operating pressure is about 220 psi during oxygen-blown gasification.

For start-up purposes, a direct propane-fired burner is provided at the gasifier mixing zone. Coal and sorbent (when required for sulfur capture) are separately fed into the Transport Gasifier through lockhoppers. Coal is ground to a nominal particle diameter between 250 and 400 microns. Sorbent, either limestone or dolomite, is ground to a nominal particle diameter of 10 to 100 microns.

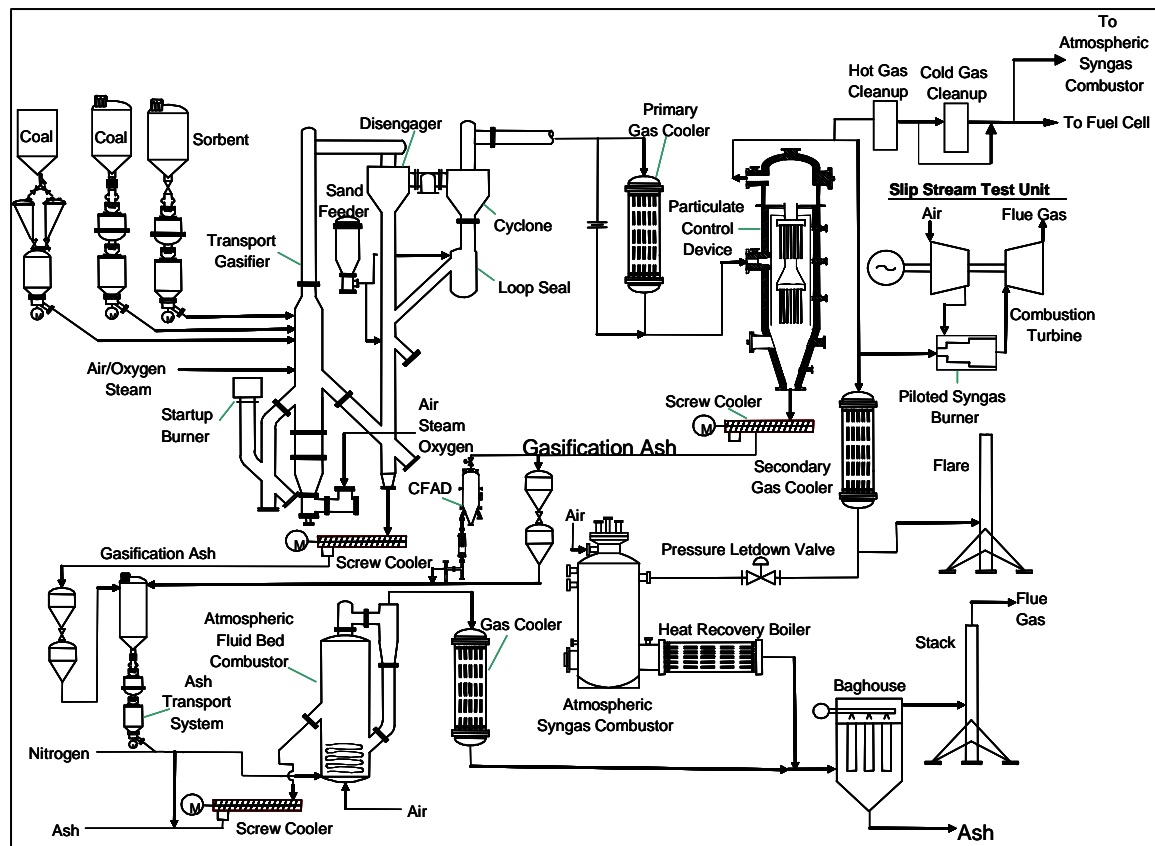
The gas exits the Transport Gasifier cyclone and goes to the primary gas cooler and then to the PCD for final particulate clean-up. The metal or ceramic filter elements used in the PCD remove

almost all the dust from the gas stream, preventing erosion of downstream equipment and controlling particulate emissions from the plant. Shown in Figure 2.1-3, the PCD utilizes a tube sheet holding up to 91 filter elements, which are attached to one of two plenums. Process gas flows into the PCD through a tangential entrance, around a shroud, and through the filter elements into the plenums. Failsafe devices are located on the clean side of the filter elements to stop solids leakage in the event of element failures. High pressure nitrogen backpulsing, typically lasting 0.2 seconds, is used to remove the accumulated solids and control the pressure drop across the tube sheet. The solids fall to the PCD cone and are removed through a lock hopper system or the continuous fine ash depressurization system (CFAD).

After exiting the PCD, a portion of the syngas can flow to the piloted syngas burner (PSB), where the gas is combusted using air from the turbine compressor. Propane supplied to the PSB serves as a pilot for the burner, as well as, a supplement to the syngas fuel to maintain burner flame stability. After combusting in the burner, the gas passes through the turbine before exiting the turbine stack. An associated generator supplies power to the electricity transmission grid. The PSB and turbine system are capable of running independently of the gasifier by using propane alone as fuel. A small portion of the syngas can also flow to a specialized gas cleanup system downstream of the PCD. The gas cleanup system removes sulfur, nitrogen, and chlorine compounds, providing a syngas suitable for use in a fuel cell.

The syngas not flowing to the PSB or cleanup system continues to the secondary gas cooler and then passes through a pressure control valve. The gas is then sent to the atmospheric syngas combustor (thermal oxidizer) which oxidizes carbon monoxide, reduced sulfur compounds (H_2S , COS , and CS_2), and reduced nitrogen compounds (NH_3 and HCN). The atmospheric syngas combustor uses propane as a supplemental fuel. The gas from the atmospheric syngas combustor goes to the heat recovery boiler, through the baghouse, and then to the stack.

The Transport Gasifier produces both fine solids filtered by the PCD and coarse solids extracted from the gasifier standpipe. The two solid streams are cooled using screw coolers, reduced in pressure in lock hoppers and then combined together. The CFAD system is also available for removing fine solids from the PCD. Coarse and fine solids removed from the gasifier and the PCD are combined, sent to the atmospheric fluidized bed combustor, cooled, and sent for disposal.



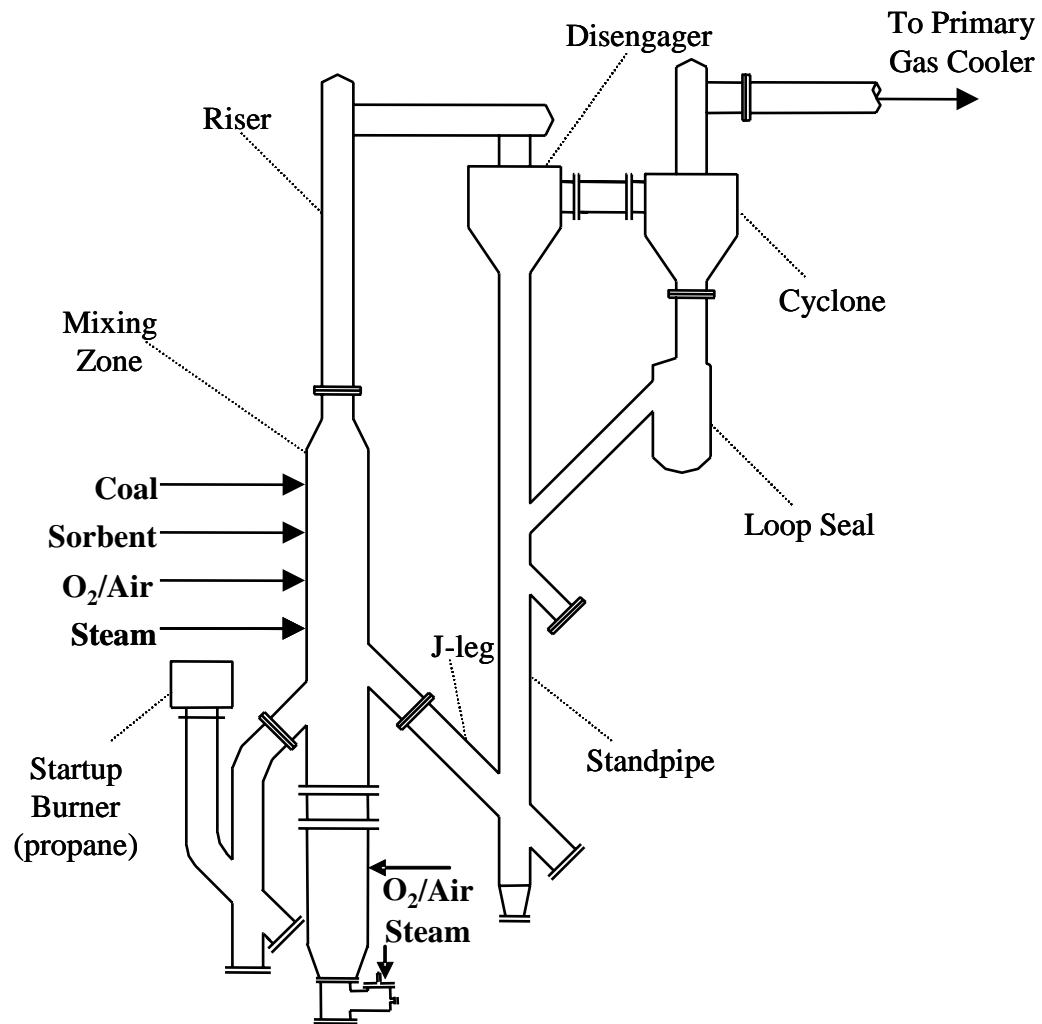


Figure 2.1-2 KBR Transport Gasifier

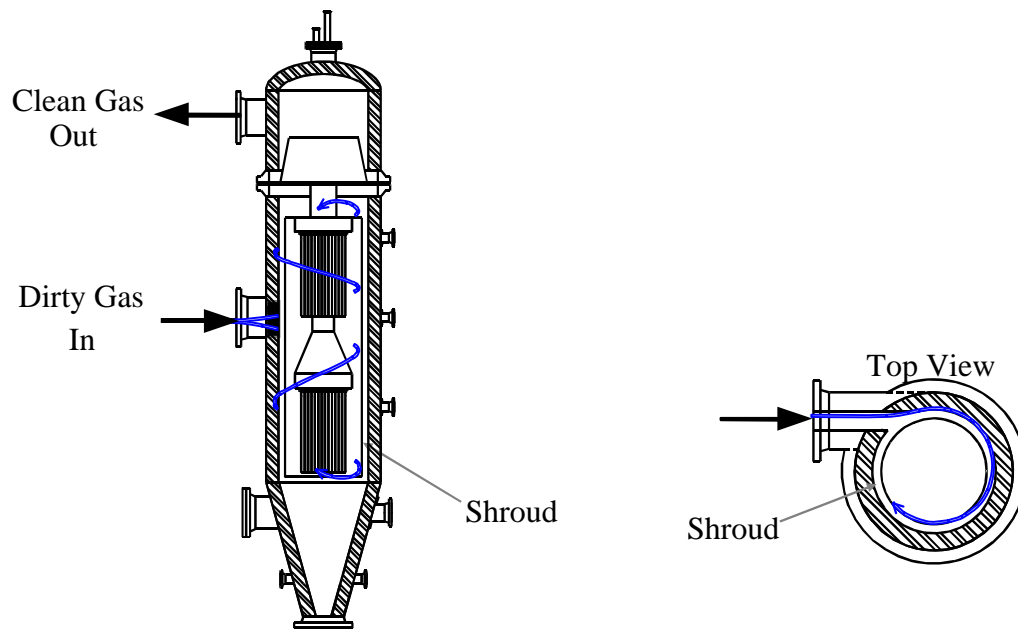


Figure 2.1-3 Siemens Particulate Control Device

2.2 DETAILED TEST CAMPAIGN OBJECTIVES

TC16 was planned as a nominally 1000-hour test run to evaluate gasifier and particulate control device (PCD) operations using Powder River Basin (PRB) subbituminous and high sodium lignite coal. The primary test campaign objectives were:

- **Fuel Cell Testing** – The overall objective for the fuel cell testing was to demonstrate reliable operation using coal derived syngas while operating with PRB coal in oxygen blown mode. This was accomplished during TC16 by operating a Delphi solid oxide fuel cell for 118 hours. The first fuel cell stack ran for 28 hours, during which time the performance declined significantly due to a high concentration of high molecular weight hydrocarbons in the syngas. As a result, a new hydrocarbon cracker was installed in the hot gas cleanup train for testing the second fuel cell stack. The new hydrocarbon cracker slowed the degradation, and the fuel cell performance was stable for 82 hours. The second test could have continued for a longer period of time; however, the test stand control system malfunctioned resulting in an earlier system shut down than was anticipated.
- **Testing of Higher Pressure Oxygen Blown Operations** – The new steam/oxygen eductor was commissioned, allowing higher oxygen/steam mixture supply pressures. The eductor operated stably, blending the steam and oxygen and allowing oxygen addition at higher gasifier pressures. The eductor system boosted the oxygen pressure to support operations at gasifier pressures up to 215 psig in the LMZ, whereas with the original oxygen/steam mixer limited gasifier pressure to a range up to 155 psig.
- **High Sodium Lignite Operations** – High sodium lignite testing was performed to characterize Transport Gasifier and PCD operation. The Transport Gasifier was successfully operated for 182 hours with high sodium lignite from the Freedom mine in North Dakota in air-blown mode. During the lignite portion of the test run, the size and shape of the standpipe particles did not change over time, indicating that no agglomerations were forming. PCD operations were stable throughout the run. PCD temperatures and pressure drop were stable while testing lignite. Bridging did not occur during the lignite portion of the run.
- **Failsafe Testing** – During TC16, a new system utilizing rupture discs was used to characterize the Pall fuse performance under catastrophic filter element failure conditions. The new system performed well. After establishing a leak-free baseline (less than 0.1 ppmw), a series of PCD outlet sampling runs were conducted in conjunction with the failsafe testing. The outlet samples indicated the failsafe performed well, and the outlet concentration tested approximately 4 hours after exposure was below the detection limit.
- **Backpulse Pressure Characterization** – Three pressure taps were installed to measure pressure changes inside a filter element, inside the plenum, and in the PCD vessel. To develop further understanding of baseline parameter, measurements with a high speed data acquisition device were taken at various operating conditions to evaluate backpulse intensity.
- **Syngas Cleaning for Fuel Cell Operation** – Bulk cleaning and polishing syngas in hot and cold gas cleanup streams using various sorbents, catalysts, and chemical solutions was used to provide clean syngas for fuel cell testing. Both the hot and cold gas cleanup units ran well without plugging. The gas cleanup units were able to supply the fuel cell with syngas containing less than 100 ppb sulfur compounds and less than 1 ppm HCl.

- **Gas Sampling System** – The new gas analyzer systems (FTIR and GC with a flame photometric detector) were commissioned. Improvements in gas conditioning techniques provided consistent gas samples to the analyzers. The operational reliability and measurement accuracy of these analyzers were demonstrated throughout the testing.

Secondary objectives included the following:

- **Oxygen Distribution System Testing** – The new steam, oxygen, and air/nitrogen flow control valves and measurement devices installed on the oxygen/steam supply system to the UMZ were tested. The air and steam flow control valves and measurement devices were fully commissioned and performed as designed. Additional testing was successful in distributing the oxygen flow to the UMZ and verifying the oxygen flow measurement. Adding oxygen into the UMZ resulted in a lower temperature drop from the UMZ to the gasifier exit.
- **Continuous Fine Ash Depressurization (CFAD) System Development** – Testing of the CFAD system continued. The CFAD system worked well and ran for 834 hours during TC16. The PCD solids carryover rate was high enough during most of the test campaign to allow the CFAD unit to operate in continuous mode. The new automatic level control was tested at the end of the test run and worked well.
- **Piloted Syngas Burner (PSB) Testing** – The PSB operated on syngas for around seven hours, at syngas flow rates up to 13,000 pph, and the combustion turbine operated for about 20 hours during TC16.
- **Pressurized Sand Feeder Commissioning**– In attempt to improve startup operations, the new pressurized sand feeder system was successfully commissioned and was able to add sand to the gasifier at pressures up to 60 psig.
- **Sensor Development/Automation** – To improve gasifier instrument longevity, the study of various materials of construction and installation methods of thermowells and pressure differential instruments continued. The evaluation of nuclear density solids flow instrumentation was also continued. Parametric tests were conducted to gather data for the gasifier temperature profile predictive model. Horizontal thermowells extending one inch past the mixing zone center line did not suffer the effects of stem loss temperature errors; however, thermowells only penetrating two inches past the refractory yielded measurement errors in excess of 30 degrees.
- **Annual Air Compliance Stack Testing** – The annual air compliance stack testing was performed by an independent environmental testing firm and the results submitted to the Alabama Department of Environmental Management. Testing was performed on the KBR stack and on the Foster Wheeler stack with the PSB fired on syngas. The test results confirmed the facility is in compliance with state regulation and its air permit.
- **Optimization of PCD Backpulse Parameters** – During the run, varying backpulse frequencies and pressures were utilized, and the pressure drop response and filter element cleaning was monitored. During the July portion of the run, the top and bottom backpulse pressures were initially kept at 150 psi above system pressure, while the backpulse cycle time was varied from 5 to 20 minutes. These relatively less stringent parameters were employed in an effort to establish the lower limit of backpulse parameters needed to maintain adequate filter element cleaning. Because bridging was eventually indicated by process characteristics and performance, these lower limits were bound.

- **Particulate Characterization and Drag Evaluation** – Gasification ash samples were collected in situ at the PCD inlet to thoroughly characterize and examine the effects of changes in coal properties, carbon conversion, sorbent addition, and operating conditions. The characterization included measurements of particulate loading, particle size distribution, chemical composition, physical properties, and drag. Measurements made in the laboratory showed good agreement with the transient drag values determined from the PCD pressure drop and particulate loadings.
- **Particulate Collection Performance** – Particle concentration was measured at the outlet of the PCD throughout TC16, and the collection efficiency of the PCD was determined to characterize PCD performance. A small amount of particle penetration of up to 0.4 ppmw was indicated during the first two days, but dropped below 0.1 ppmw for the remainder of the test run.
- **Turbine Exhaust Sampling** – A single particulate sampling run was performed in the stack during operation of the combustion turbine in simple-cycle mode (propane firing with no syngas). The sampling filter was contaminated with a large amount of material that appeared to be rust scale or deposits from the stack itself. The presence of this foreign material made it difficult to quantify the amount of ash present during operation on syngas, and the subsequent period of operation on syngas was not sufficient to perform another sampling run. Despite these problems, the sampling in the stack did verify that no soot was present, suggesting good operation of the combustion turbine on propane.
- **Ongoing Evaluation of PCME Particulate Monitor** – Throughout TC16, the response of the PCME particulate monitor was evaluated to gain a better understanding of the instrument response and to detect any particle penetration through the PCD. Dust was injected into the PCD outlet to stimulate the monitor and define its response. The PCME response was also monitored during the failsafe test, and no high level emissions were detected.
- **Commissioning of Minireactor Systems** – The minireactor system was commissioned in two phases utilizing a Sud-Chemie catalyst for ammonia and hydrocarbon cracking over a period of 500 hours with gasifier syngas. In the first phase, a minireactor of 1.5 inch ID was used for the syngas slip stream cleanup, while the second phase used a new minireactor of 1.687 inch ID to support fuel cell testing. The hydrocarbon content of the syngas was significantly reduced by catalytic cracking.
- **Gas Cleanup Process Performance** – Various methods of optimizing gas cleanup performance were performed throughout TC16. The effects of operating parameters such as space velocity, bed height and temperature on sorbent sulfur capturing capacity, exit sulfur concentration level, sulfur break-through time, sorbent strength and elutriation, ammonia and hydrocarbon cracking continued to be evaluated throughout the testing. During the last 72 hours of fuel cell operation, the gas cleanup system was able to reduce the total condensable organic content in the syngas to below 300 ppm by cracking the organics using the nickel-based catalyst at 1,600°F.

2.3 DETAILED TEST CAMPAIGN SUMMARY

The activities that occurred during the outage preceding test run TC16 included 40 equipment modifications. The most significant of these are listed below:

- An eductor was installed in parallel to the steam/oxygen mixer to boost the mixture supply pressure so that the gasifier could be operated at a higher pressure during oxygen-blown mode. The new steam/oxygen eductor operated very well, blending the steam and oxygen and allowing oxygen addition at a gasifier pressure of 215 psig in the lower mixing zone (LMZ).
- Selected horizontal thermowells in the mixing zone were lengthened for evaluating the effects of stem losses and thermal shunting. Thermowells extending one inch past centerline did not suffer the effects of stem loss temperature errors as the bed material transitioned from dense to dilute phase. The shorter thermowells, penetrating only 2 inches past the refractory, yielded measurement errors in excess of 30°.
- Three robust thermowells with a 1.5 inch diameter were installed at a relatively steep 30° angle to the gasifier shell. These thermowells were installed to penetrate past the centerline and were added at different elevations in the riser. The thermowells were made of Hastelloy X for coefficient of expansion compatibility with the Pyrocil D sheath material of the 0.375 inch diameter type N thermocouple. The Hastelloy X was coated with approximately one-eighth inch of Stellite # 1 using a plasma transfer process. The penetration of these thermowells into the process provided baseline measurements since stem loss effects and thermal shunting were negligible.
- Separate steam, oxygen, and air/nitrogen flow control valves and measurement devices were installed on the oxygen/steam steam supply system to the UMZ. The air and steam flow control valves and measurement devices were fully commissioned and operated as expected. Additional testing was successful in distributing the oxygen flow to the UMZ and verifying the oxygen flow measurement. Adding oxygen into the UMZ resulted in a lower temperature drop from the UMZ to the gasifier exit.
- A lower heating value sensor supplied by General Electric (GE) was added to the syngas cleanup system for testing. GE support performed several gas analysis tests during oxygen-blown operations, during a one hour transition from oxygen- to air-blown mode, and for an additional 5 hours in air-blown mode.
- A vibrator was installed on the developmental coal feeder lock vessel to promote coal dropping into the feeder. There did not appear to be any operational issues with the developmental coal feeder related to coal dropping into the feed vessel.
- In preparation for fuel cell testing, the hot gas cleanup system was modified, and the syngas cleanup units were operated for a total of 577 hours. Both the hot and cold gas cleanup units ran well without plugging. Improvements in gas conditioning techniques provided consistent gas samples to the analyzers at the gas cleanup skid. The gas cleanup units were able to supply the fuel cell with clean syngas that contained less than 100 ppb sulfur compounds and less than 1 ppm HCl.

Typical operating parameters for the Transport Gasifier and the PCD during TC16 are shown in Table 2.3-1. Operating trends for the run can be seen in Appendix 4, Figures A4-1 through A4-29. The following is a brief operating synopsis of TC16:

Startup Activities

PCD preheat began with a gasifier pressure of 60 psig. The new pressurized sand feeder was operated to add sand to the gasifier while at pressure. Once the PCD temperature was above 200°F for over 1 hour, the test run commenced with the lighting of the startup burner on July 13, 2004. Motor problems on the atmospheric syngas burner fan precluded coal feed.

July 13, 2004

The standpipe screw cooler was run, but it depleted the gasifier solids at a rate faster than desired, so it was shut off after a few seconds. The lock-hopper fines removal system experienced problems with several valves leaking which prevented adequate operation. Thus, CFAD was operated exclusively for spent fines removal. Preliminary pressure measurements were conducted on the PCD failsafe tester to characterize the new system.

July 14, 2004

Riser velocity was reduced to stem the loss of solids level in the gasifier standpipe. Once the riser velocity decreased, coal feed was initiated, and the burner firing rate was reduced. The gasifier pressure in the UMZ was around 125 psig, and the riser velocity was about 50 ft/s.

Throughout the night, the coal feeder tripped due to a logic problem. A small adjustment was made in the programmable logic controller (PLC) to prevent additional avoidable trips, and the feeder began operating normally. When conditions stabilized, the startup burner was shut down, and the gasifier temperatures were slowly increased by feeding coal (via the FD0210 coal feeder) at a low feed rate. When the gasifier temperature increased above 1,650°F, the coal feed rate was increased, and syngas quality improved dramatically.

Steam addition was started at various locations in the gasifier as the temperatures increased further.

July 15

The highest gasifier temperature of 1,825°F was in the UMZ, while the gasifier exit temperature was just below 1,650°F. During the day, the gasifier began to continuously lose solids at a rate of about 20 pph. Sand was added to increase the standpipe level. However, the level still gradually decreased after the sand was added. Other than the loss of solids, the gasifier performed well. The new upper oxygen distribution nozzles were initially commissioned with air to evaluate the control valves and instrumentation. The two flow meters appeared to read correctly.

Since the PCD differential pressure (DP) was low, the backpulse timer was increased to 15 minutes. Shortly thereafter, backpulse tests were performed for the standard 0.2 second valve open time. The backpulse pressures were 150, 200, 250, and 300 psi above system pressure during testing. Pressure measurements were taken at three locations to evaluate the backpulse intensity across filter elements at these operating conditions. When the tests were complete, the backpulse pressure was lowered to 150 psi above system pressure.

The gasifier continued to lose bed material at a variable rate throughout the night, so sand was added periodically. Due to the high feed rate of ambient temperature sand, the gasifier temperatures decreased significantly during sand addition. However, the system recovered quickly, and operated well for the remainder of the night. The steam and air flow to the LMZ were increased and the flows to the new air/oxygen nozzles in the UMZ were decreased.

July 16, 2004

In preparation for transitioning to oxygen-blown mode, the gasifier pressure was reduced. The air flow rates, coal feed rate, and aeration rates were also reduced to maintain a proper riser velocity. The oxygen transition was successful, and gasifier operations were smooth; however, the excessive loss of bed material continued. A backpulse test was conducted on the PCD for a 0.4 second valve open time. The baseline pressure drop decreased slightly compared to the normally used 0.2 second valve open time.

July 17, 2004

Material loss from the gasifier continued. Various aeration adjustments were made in the gasifier loop but did not appear to have an apparent effect on the loss rate. Periodic sand addition continued to maintain bed levels in the gasifier.

During the day, syngas flow to the gas cleanup processes began so that the gas analysis systems could be evaluated in preparation for sending syngas to the fuel cell. Since the PCD peak pressure drop was consistently low and stable, the backpulse timer was increased to 20 minutes.

The gasifier operated steadily through the night. All the aeration associated with the cyclone dipleg was stopped as it had no significant effect on the cyclone performance. Adding sand to the gasifier was still necessary at times to maintain the standpipe level. The fuel cell preheat was initiated.

July 18, 2004

System operation continued to be smooth throughout the day and night. Syngas flow to the fuel cell was started.

July 19

Gasifier operating parameters were adjusted to evaluate their effects on the rate of bed material loss. Limestone was added, and the standpipe level remained constant for a short period of time; however, it later began declining again. Some responses from the PCD resistance probes indicated gasification ash bridging, even though the baseline pressure drop and thermocouple responses did not indicate any problems.

July 20, 2004

The fuel cell was shut down in the afternoon due to a problem with the fuel cell control system. About 150 pph of limestone was added throughout the day to maintain the standpipe level.

Parametric testing on the PCD was performed with different backpulse pressures to evaluate the effect of a higher backpulse pressure on the resistance probe measurements. The higher pressure appeared to lower the resistance probe readings. Therefore, the backpulse pressure was lowered to 150 psi above gasifier pressure on both plenums, and the valve open time was changed to 0.5 seconds.

July 21, 2004

The system was steady throughout the day and night, but the loss of bed material continued. The limestone feed was continued at a rate of around 170 pph to maintain the standpipe level. The resistance probes in the PCD continued to show signs of gasification ash bridging. Some PCD thermocouples were less responsive than others and did not trend as closely as normal to the other thermocouples, also indicating bridging.

July 22, 2004

The gasifier inventory increased gradually throughout the day due to higher coal particle size and limestone addition. The CFAD system also operated well with a solids removal rate of around 800 to 1000 pph, well above the actual PCD solids carryover rate of approximately 400 pph. The higher PCD resistance probes readings continued but were stable. A solids injection test was performed at the PCD outlet to evaluate the response of the PCME particulate monitor.

Gasifier operations were smooth, and the standpipe level increased some after sand addition. The coal feed was reduced, and a nearly constant standpipe level and solids circulation rate was maintained. The limestone feed testing for TC16 was stopped.

July 23, 2004

Dolomite addition to the gasifier was started mid-morning, but was later stopped to prevent too high a standpipe level. PCME calibration testing continued. The fuel cell operated on calibration hydrogen throughout the early morning.

July 24, 2004

The gasifier operated smoothly throughout the day. Dolomite addition was restarted to prepare for the lignite run and minimize the amount of sand in the circulating solids.

The hot gas cleanup minireactor heat-up was started, and syngas flow was later initiated. Early measurements showed that the mini-reactor was cracking the hydrocarbons, converting over 95 percent of the benzene and over 70 percent of the methane. Syngas flow to the fuel cell was initiated that afternoon and continued throughout the night.

The standpipe level increased gradually throughout the night, so the screw cooler was operated to reduce the level.

July 25, 2004

Gasifier operations continued smoothly with dolomite addition. The standpipe solids removal system was operated as needed to maintain an acceptable solids level. The fuel cell ran well, generating around 550 W at 22 V. The fuel cell tripped a few times during the night due to software communication problems.

July 26, 2004

The gasifier solids removal system operated periodically to maintain a constant standpipe solids level. The gas cleanup system and fuel cell operated continuously throughout the day. Minor adjustments to the AFBC operating parameters optimized its performance.

The gasifier tripped during the night when the coal feeder experienced difficulty transferring fine coal from the surge bin to the dispense vessel. Minor temperature spikes were seen in the PCD in which some thermocouples increased by 50 to 100°F. Eventually, the process was restored to steady state oxygen-blown operations.

July 27, 2004

The coal and oxygen flow rates were reduced while the coal feeder vent lines were unplugged. The gasifier returned to steady state operations after the problem was resolved. A few differential pressure ports plugged around midnight, and some of the thermocouples showed signs of failing.

July 28, 2004

Gasifier operations continued to be stable, and the fuel cell testing was completed. The PSB inlet line was preheated in preparation for testing scheduled for the next morning.

July 29, 2004

The coarse solids screw cooler experienced some operational problems. Solids circulation was disturbed momentarily as adjustments were made to the lower standpipe and screw cooler purge aeration flows. The PSB inlet line preheating continued, while simple cycle testing was performed on the CT. Several gas analysis tests were performed by GE to evaluate the performance of the lower heating value sensor.

July 30, 2004

GE completed their testing in oxygen-blown mode, and the gasifier was transitioned back to air-blown mode. GE continued testing during the one hour transition and for five hours in air-blown mode. Syngas was initiated to the PSB/CT to test the controls in preparation for the turbine stack testing. Dolomite feed was stopped.

July 31, 2004

Due to continued operational problems with the coarse solids screw cooler, the coal feed rate was reduced to minimize the accumulation of solids in the gasifier.

August 1, 2004

Gasifier operations were stable with a slow increase in the standpipe level. The coal feed rate was reduced further to evaluate the standpipe level rate of increase. The low coal feed rate minimized the accumulation of solids in the gasifier. Annual air compliance emission tests on the PSB/turbine and the MWK stack were performed. Syngas flowed to the gas turbine for about four hours at a flow rate of 12,000 pph. The turbine output was 1.6 MWe with a propane flow rate of 700 pph.

August 2, 2004

The standpipe level was too high, but solids could not be removed from the standpipe screw cooler because it was out of service. Other attempts to remove solids resulted in solids circulation upsets.

August 3, 2004

To prepare for failsafe testing, the PCD backpulse pressure was changed to 150 psi above system pressure on both plenums, the cycle time was set at 10 minutes and the valve open time was changed to 0.2 seconds. After the parameters were set, the failsafe test was performed. Although the rupture disk opened during the first attempt, it was not realized until some time later, and outlet sampling was not conducted during that time. After the test, the lower plenum backpulse pressure was increased to 250 psi above the reactor pressure. PCD outlet sampling showed less than 0.1 ppmw which indicated that the Pall fuse did not leak four hours after the test device opened.

Later in the day, the process was shut down and the system was depressurized to repair the standpipe screw cooler and inspect the cyclone. Immediately after coal feed was stopped, a controlled burn of the PCD filter cake was performed. Once coal feed was terminated, air flow was re-started to the gasifier to give approximately 3 percent oxygen at the PCD. After about three minutes, the filter element temperatures increased sharply, indicating that the filter cake material was being burned.

Outage: Tuesday, August 3 through Monday, August 9, 2004

A scheduled outage occurred to inspect the coarse ash screw cooler, the cyclone, and the PCD, and to repair a leak on the condensate system. When the standpipe was drained, a few pieces of refractory and deposited material were found. They were apparently left over from TC15 but did not appear to be large or hard enough to prevent the screw cooler from operating normally. Also, the cyclone was missing a piece of refractory at the gas inlet, and the notch at the cyclone inlet was reforming. Inspection of the PCD revealed some remaining bridged material.

August 9, 2004

The test run resumed, and the startup burner was lit. Gasifier temperatures slowly increased and the AFBC fuel oil injection was started.

After the startup burner heated the gasifier to around 1,100°F, coal feed began. Just before coal feed began, steam flow to the LMZ and coal steam shrouds was started.

August 10, 2004

In preparation for oxygen-blown testing, the coal feed rate, the steam flow rate, and the system pressure were increased. The gasifier was transitioned to oxygen-blown mode operation at a coal feed rate of about 3,000 pph. Operations were smooth. Further tuning of the oxygen flow rate controller was performed, as well as lining the system up to commission the new eductor. Transitioning to the eductor was smooth. Since the standpipe level was decreasing, the coal feed rate was increased. The increase in coal feed rate caused the standpipe level to increase slowly. The eductor operating envelope was developed by varying the steam and oxygen flows. The flows were adjusted to minimize the temperature swings. PCD Backpulse pressure testing was performed, and the backpulse valve open time was adjusted to 0.4 seconds from 0.5 seconds.

August 11, 2004

The LMZ pressure was increased to 200 psig. The transition was smooth other than the maximum pressure setting (195 psig) in the logic tripped the oxygen valve closed twice. The logic was modified, and operations were smooth throughout the day at an operating pressure of 200-205 psig.

The FTIR analyzers were online periodically to measure hydrocarbon concentrations. The gasifier continued to run smoothly at the higher pressure oxygen-blown conditions.

August 12, 2004

The main air compressor tripped, causing the entire system to trip. The gasifier was restarted and quickly transitioned back to oxygen-blown mode. The gasifier performed well throughout the day at a LMZ pressure of 200 psig. The effect of the gasifier exit temperatures on the hydrocarbon species was tested. The benzene concentration decreased from 1,400 to 1,000 ppm as the exit temperature increased from 1,650 to 1,700°F. The oxygen and steam flow control valves were tuned. Oxygen flow was initiated into the gasifier through the UMZ nozzles.

August 13, 2004

The eductor steam flow controller was tuned. The oxygen flow to the UMZ was increased, resulting in a lower temperature drop from the UMZ to the gasifier exit. Problems with the coal feed system developed. The surge bin was emptied and the feeder operations improved. The system was transitioned back to air-blown operations. Temperatures in the PCD increased after the transition.

August 14, 2004

The developmental coal feeder was used to feed coal to the gasifier. Automation parametric tests took place throughout the day. Parametric testing was interrupted when the developmental coal feeder plugged. The feeder was unplugged, and the coal feed rate from the original coal feeder was increased. Fines again caused operational problems with the original coal feed system, so the coal feed rate was reduced while the material was fed. Once the fines were emptied from the surge bin, the original feeder worked well, and the gasifier resumed normal operations.

August 15, 2004

Some controls improvements were made while the developmental coal feeder was operating in offline mode. The developmental coal feeder was then lined up to feed coal to the gasifier, and the coal feed rate through the original coal feeder was reduced. Operations were smooth except, with the exception of one sudden decrease in coal feed, which resulted in temperature swings in the gasifier. Additional parametric tests were performed. When the testing was completed, the developmental coal feeder was shut down, and the coal feed rate was increased on the original coal feeder. Shortly thereafter, there were some minor problems with coal flow from the surge bin to the lock hopper in the original coal feeder which caused one brief trip. The process was quickly returned to stable operations.

August 16, 2004

The developmental coal feeder was started around noon. Other than a few minor problems, the feed rate was smooth. The standpipe level control was placed in automatic since the gasifier was continually increasing in solids inventory. The developmental coal feeder continued to have problems with the discharge line plugging, so it was transitioned to offline mode to further evaluate its operation.

August 17, 2004

The gasifier pressure and temperature were reduced simultaneously with the coal feed rate in preparation for feeding high sodium lignite. When the gasifier temperatures reached 1,550°F, lignite feed to the gasifier was initiated. The transition was smooth except for a minor plugging of the original coal feeder conveying line. To evaluate gasifier operations with lignite, loop seal, and standpipe solid samples were closely monitored to detect any agglomerations. The spent fines screw cooler developed a small leak but was repaired.

August 18, 2004

Gasifier and PCD operations were stable. The dolomite feed rate was increased. The loop seal and standpipe solid samples appeared to have no agglomerations present.

August 19, 2004

The gasifier and PCD continued to operate well with virtually no changes in operation.

August 20, 2004

The loop seal and standpipe solid samples continued to look acceptable with no agglomerations present. The CFAD system continued to operate well and some initial work was performed to automate the level, the CFAD level control.

August 21, 2004

Gasifier and PCD operations were stable.

August 22, 2004

The gasifier temperature was decreased slightly due to a decrease in standpipe level. Within a few hours the standpipe level increased, and the standpipe level control was placed back in automatic control. A leak developed on the spent fines screw cooler during the night. Attempts were made to repair but met with only limited success.

August 23, 2004

The backpulse timer was changed to 4, 5 and 7 minutes in order to perform CFAD level control tests. The tests were conducted to verify that the CFAD level indicator was responding directly to the PCD backpulse.

August 24, 2004

After all the lignite was fed, coal feed was stopped, giving a total of 835 hours of coal feed for TC16. A controlled burn procedure was performed on the filter elements. Gas with approximately 3 percent oxygen was added to the PCD to test controlled dust cake combustion. After about 15 minutes, the temperatures increased by 300°F. When the tests were completed, the remaining systems were shut down and the gasifier was emptied of solids.

Table 2.3-1
Typical Operating Conditions for the Transport Gasifier and Particulate Collection Device

Transport Gasifier	
Startup Bed Material	Sand, ~120 μm
Startup Fuel	Coal (PRB)
Fuel Type	PRB Freedom Lignite
Fuel Particle Size (mmd), μm	142 – 533 (PRB) 191 – 378 (Lignite)
Average Fuel Feed Rate, pph	2,400 – 4,500
Sorbent Type	Ohio Bucyrus limestone Plum Run Dolomite
Gasifier Temperature, $^{\circ}\text{F}$	1760 – 1850 (PRB) 1500 – 1600 (Lignite)
Gasifier Exit Pressure, psig	141 – 191 (O_2), 217 – 226 (air) (PRB) 150 – 160 (air) (Lignite)
Riser Gas Velocity, ft/s	40 – 55 (PRB) 45 – 50 (Lignite)
Standpipe Level, inH ₂ O	80 – 160
Total Gas Flow Rate, pph	13,000 – 24,000
Air/coal mass ratio	2.4 – 3.4
Oxygen/coal mass ratio	0.5 – 0.8
Oxygen/steam mass ratio	0.4 – 0.9
Steam/coal mass ratio	0.2 – 1.3 (PRB) 0.3 – 0.9 (Lignite)
Particulate Control Device	
PCD Temperature, $^{\circ}\text{F}$	650 – 800
PCD Inlet Loading, ppmw	9,600 – 44,300
PCD Outlet Loading, ppmw	< 0.1
PCD Baseline Pressure Drop, inH ₂ O	60 – 120
Type of Filter Elements	Pall Iron Aluminide
Filtration Area, ft ²	204.5
Face Velocity, ft/min	3.0 – 4.0
Pulse Valve Open Time, sec	0.2 – 0.5
Pulse Cycle Time, min	5 – 20
Pulse Pressure, Top Plenum	150 – 250 psi above System Pressure
Pulse Pressure, Bottom Plenum	150 – 250 psi above System Pressure

2.4 INSPECTIONS

2.4.1 Transport Gasifier

The Transport Gasifier performed well throughout the test campaign. During the inspections, the gasifier was found to be relatively clean and in good condition. There were few deposits found in the gasifier loop. However, the notch in the cyclone inlet appeared to be redeveloping as seen in Figure 2.4-1.

The riser and the LMZ were both clear. The mixing zone was coated in deposits very similar to those in the loop seal downcomer, which had not been seen before. The deposits were very thin and were easily removed. The secondary crossover was remarkably clean, but the riser crossover had the usual large, loose pile of material.

The disengager appeared to be in good condition, but there were a few small deposits along the wall in some places. The cyclone refractory was in decent condition overall with the exception of the notch that redeveloped in the inlet. Additional inspections will be performed after the next run. There was also a continuous but thin coating of deposits in the cyclone which were more severe in the cone section. The loop seal had the usual small, soft, bumpy deposits on the walls of the downcomer. The deposits appeared to be thicker than in the past, but flaked off easily when touched by the boroscope.

The standpipe was clean and the screw flights on the coarse solids screw cooler (FD0206) were visible.

The primary gas cooler tubesheet was in fairly good condition, with only a slight coating of dust and some small pieces of material in a few ferrules.

2.4.2 Particulate Control Device

The PCD was inspected twice. The first inspection occurred during an outage in the middle of the test run. The lower manway was opened, and the bottom plenum was inspected. Bridged material remaining after the dustcake burnoff was present on a section of filter elements that covered approximately one-fourth of the bottom plenum surface. Figure 2.4-2 shows the bottom plenum with the section of bridged gasification ash. The remaining bridged gasification ash was physically removed to the extent possible. The top plenum was viewed through a nozzle, and no gasification ash bridging was seen. Some gasification ash was found in the pressure taps on the nozzles when they were disconnected. Apparently, the purge was not strong enough or the gasification ash penetrated into the tubing while pressure measurements were made.

An inspection performed after the final shutdown revealed one broken iron aluminide filter on the bottom plenum. The filter was broken at a welded junction, possibly due to the bridged gasification ash in the first portion of the test run. The Pall fuse failsafe above the filter element was plugged. This failsafe was very effective in preventing particle penetration since the outlet loading samples were below 0.1 ppmw for the majority of the test run. A very heavy dustcake was observed on the top portion of the filter elements. The dustcake on the bottom portion of the filter elements was not as heavy as the top.

Fifteen of the filter elements were removed and flow tested using air at ambient temperature and pressure. The results revealed that six filter elements had pressure drops between 4 and 12 inH₂O at a face velocity of 3 ft/min. These filter elements were acceptable for reinstallation in later test campaigns. The remaining nine filter elements had pressure drops between 22 and 34 in H₂O at a face velocity of 3 ft/min, which was too high for reinstallation.



Figure 2.4-1 Notch at Entrance to Cyclone

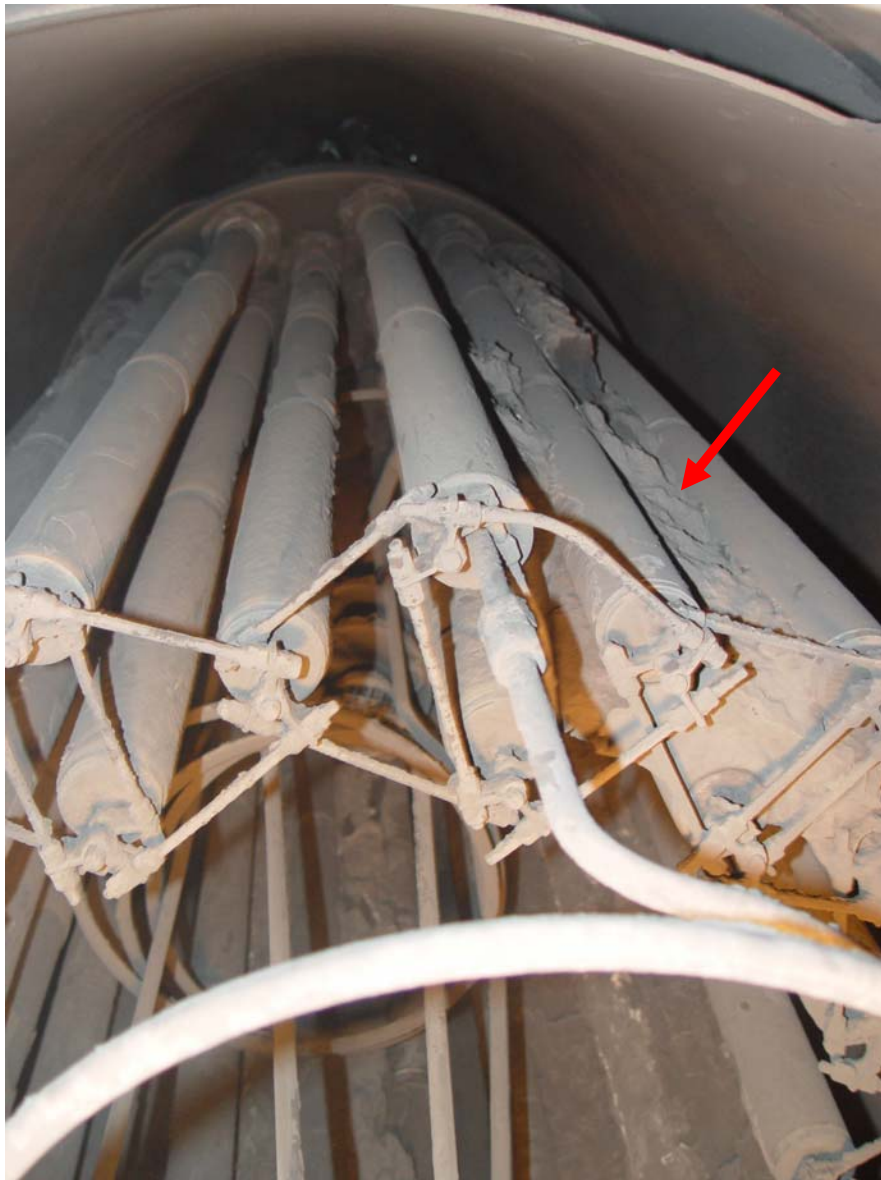


Figure 2.4-2 Bridged Gasification Ash on Bottom Plenum of PCD

3.0 PERFORMANCE

3.1 TRANSPORT GASIFIER PERFORMANCE

3.1.1 Overview

The test campaign consisted of 55 periods of steady operation between July 15 and August 24, 2004. These periods are given in Table 3.1-1. The operating periods had a cumulative time of about 368 hours, which was about 44 percent of the total TC16 on-coal operation time. The first two periods occurred during air-blown operation with PRB coal. The next 21 periods were during oxygen-blown operation with PRB coal. The next 10 periods were during air-blown operation with PRB coal. Then, there were six additional periods in oxygen operation with PRB coal, followed by three periods in air blown with PRB coal. The final 13 periods were air blown with lignite coal. Ohio Bucyrus limestone and Plum Run dolomite were used as sorbent during TC16.

Table 3.1-2 lists the TC16 operating conditions, including the riser exit temperature, the system pressure, the PCD inlet temperature, the PCD solids rate, air rate, oxygen rate, coal rate, sorbent (limestone or dolomite) rate, syngas rate, steam rate, and nitrogen rate of the steady state operating periods. The system pressure ranged from 217 to 226 psig for the air-blown periods with PRB coal, but the pressure was reduced to between 150 and 159 psig during air blown with lignite. During oxygen-blown mode, the pressure was between 145 and 191 psig. Gasifier temperatures were between 1,648 and 1,712°F while using PRB coal, and ranged from 1,434 to 1,481°F with lignite. Steam flow rates were between 500 and 1,800 pph in air-blown mode and between 3,000 and 4,000 pph in oxygen-blown mode.

3.1.2 Gas Composition

During TC16, the Transport Gasifier and syngas combustor outlet gas analyzers were continuously monitored and recorded by the Plant Information system. Twenty-four in situ samples of syngas were taken during PCD outlet loading sampling and measured for moisture content. The locations of the gas analyzers are shown in Figure 3.3-1 of the TC12 Topical Report. Plotted on Figure 3.1-1 is the syngas moisture analyzer (AI475H) data, in situ data, and the moisture content based on a correlation of temperature and the in situ data. The moisture content during air-blown gasification averaged 10 percent for PRB coal and 17 percent for lignite. During oxygen-blown gasification, the moisture content increased to about 27 percent due to the higher steam flow rates.

The H₂O concentrations calculated for the operating periods are given in Table 3.1-3. They ranged from 8.8 to 13.9 percent and from 15.7 to 20.3 percent in air-blown gasification using PRB and lignite, respectively were between 25.7 and 30.8 percent during the oxygen-blown gasification periods. Based on these moisture concentrations, the estimated wet syngas compositions for the TC16 operating periods are given in Table 3.1-3 and shown on Figure 3.1-2. Also shown in Table 3.1-3 are the wet syngas molecular weights for each operating period.

The CO concentration typically ranged from 6.0 to 9.9 percent during air-blown gasification with PRB. The CO content was normally between 5.5 and 9.1 percent during the oxygen-blown gasification. During air-blown gasification with lignite, the CO concentration ranged from 1.9 to

3.9 percent. The CO values for lignite were lower than the values for PRB due to a lower coal feed rate and higher lignite moisture content than for PRB.

The H₂ concentration fluctuated between 5.8 and 7.8 percent during the air-blown periods with PRB coal and decreased slightly, to between 4.7 and 6.8 percent, when lignite was used. The H₂ values for lignite were lower than the values for PRB due to a lower coal feed rate. After the transition to oxygen-blown gasification, the hydrogen concentration increased to between 10.2 and 14.3 percent. The increase was due to the decrease in nitrogen dilution and the higher steam flow rates, which produced H₂ by the water-gas shift reaction.

The CO₂ concentration ranged from 8.1 to 9.2 percent during air-blown gasification with PRB coal and increased to between 9.3 and 10.1 percent with lignite. During the oxygen-blown periods, the CO₂ concentration increased to between 11.9 and 13.4 percent.

For PRB air-blown operation, the CO concentration was generally higher than the H₂ concentration, while for all the Freedom lignite the H₂ was higher than the CO. This change in concentration is due to the higher moisture content of the Freedom lignite compared to PRB. The additional moisture reacts with CO to produce more H₂ and CO₂. However, the CO concentration was lower than the H₂ concentration for TC16-24 and TC16-25 (PRB air-blown periods) due to high steam flow rates. The higher H₂ concentration at the expense of CO during the oxygen-blown testing with PRB was partially caused by the large increase of steam rate (the steam rate was approximately tripled).

During TC16, the CH₄ concentration was between 1.1 and 1.5 percent and 0.6 and 0.9 percent during air-blown operations using PRB and lignite, respectively. The methane content was between 1.3 and 2.2 percent during oxygen-blown mode.

The C₂⁺ concentration was assumed to be zero, since the gas analyzers were reading very low values.

The wet syngas molecular weight and nitrogen concentration are plotted on Figure 3.1-3. The air-blown gasification molecular weights ranged from 26.2 to 26.9 pounds per pound-mole with PRB and from 26.1 to 26.3 pounds per pound-mole with lignite coal. The oxygen-blown molecular weights were from 23.4 to 24.4 pounds per pound-mole. The decrease in molecular weights was due to higher concentrations of lower molecular compounds such as H₂, H₂O, and CO₂.

The main sulfur species in the syngas are H₂S and carbonyl sulfide (COS), and other sulfur compounds, such as CS₂ are present in small quantities. When combusted in the syngas combustor, these sulfur compounds present in the syngas are converted to SO₂. The wet H₂S concentration measured by AI419J and the calculated syngas total reduced sulfur (TRS) concentrations are listed in Table 3.1-4, as well as the sorbent feed rate and sulfur capture. The AI419 analyzers measure the gas composition on a dry basis, so the values from AI419J were corrected to allow for moisture. The syngas combustor SO₂ analyzer, AI476N, measures the total sulfur emissions exiting the system. The total reduced sulfur is an estimated value of total sulfur derived from the SO₂ content at the syngas combustor exit.

The TRS concentrations during TC16 are divided into six different cases based on the type of coal, the type of sorbent, and the gasifier operating mode (air or oxygen):

1. PRB coal, no sorbent, air-blown mode (TC16-1, TC16-2, and TC16-24 to TC16-33) – TRS emissions were 202 to 348 ppm and the average sulfur capture was 19 percent.
2. PRB coal, no sorbent, oxygen-blown mode (TC16-3 to TC16-7 and TC16-39) - TRS emissions were 379 to 457 ppm and the average sulfur capture was 6 percent.
3. PRB coal, limestone, oxygen-blown mode (TC16-8 to TC16-13) - TRS emissions were 292 to 465 ppm and the average sulfur capture was 7.5 percent.
4. PRB coal, dolomite, oxygen-blown mode (TC16-14 to TC16-23 and TC16-34 to TC16-38) - TRS emissions were 294 to 378 ppm and the average sulfur capture was 9 percent.
5. PRB coal, dolomite, air-blown mode (TC16-40 to TC16-42) - TRS emissions were 197 to 238 ppm and the average sulfur capture was 22 percent.
6. Lignite, dolomite, air-blown mode (TC16-43 to TC16-53) - TRS emissions were 733 to 962 ppm and the average sulfur capture was 17 percent.”

The TRS concentration was the highest when lignite coal was used during air-blown mode. However, it was also relatively high when PRB was used during oxygen-blown mode. Note that operating in oxygen-blown mode also increases the TRS emissions due to less TRS dilution by the nitrogen in the air. The TC16 TRS data does not show any effect of sorbent on sulfur capture as adding dolomite to PRB air-blown mode did not increase sulfur capture (cases 1 and 5) and adding limestone or dolomite to PRB oxygen-blown mode did not increase sulfur capture (cases 2, 3, and 4). There was no difference between the sulfur capture in the PRB oxygen blown with limestone and with dolomite (cases 3 and 4). The major effect on sulfur capture was the mode of operation with air-blown test periods averaging 18.5 percent sulfur capture and oxygen-blown test periods averaging 8 percent sulfur capture. This is shown graphically on Figure 3.1-5 where the “sulfur out” is plotted against the TRS sulfur concentration. Sulfur out is the amount of sulfur that exits the gasifier via the syngas and PCD fines and is expressed in ppm of sulfur in the syngas. The slope of the plot of the sulfur out and TRS concentration is the fraction sulfur remaining. For Figure 3.1-5 this slope is 0.825 (17.5 percent sulfur capture) for air blown and 0.92 (8 percent sulfur capture) for oxygen blown which are nearly the same as the average sulfur captures.

It should be noted that the H₂S analyzer AI419J was operating well below its acceptable range and is only accurate above 3,000 ppm H₂S.

The CO/CO₂ ratios calculated for each operating period are listed in Table 3.1-3. The TC16 CO/CO₂ ratio varied from 0.7 to 1.2 while using PRB coal in air-blown mode, but the ratio decreased to between 0.2 and 0.4 while using lignite in air-blown mode. The reason for the decrease in the CO/CO₂ ratio, while using lignite, was the lower gasifier operating temperature and higher steam rates for lignite. The CO/CO₂ ratio varied from 0.5 to 0.7 while in oxygen-blown mode using PRB.

3.1.3 Syngas Heating Values

Raw Syngas Heating Values. The calculated wet raw syngas LHV for each operating period is given in Table 3.1-3 and plotted on Figure 3.1-6.

The LHV is calculated using the formula:

$$\text{LHV (Btu/SCF)} = [275 \times (\text{H}_2\%) + 322 \times (\text{CO}\%) + 913 \times (\text{CH}_4\%) + 1641 \times (\text{C}_2^{+}\%)]/100$$

During air-blown gasification with PRB, the LHV ranged from 48 to 65 Btu/SCF. During air-blown gasification with lignite, the LHV typically ranged from 30 to 40 Btu/SCF. There was one exception to this during TC16-47 when the LHV dipped to 25 Btu/SCF. During this period the coal feed rate was low and the nitrogen dilution per BTU of coal fed was high. The LHV was normally between 59 and 87 Btu/SCF during the oxygen-blown periods using PRB.

Past test runs have indicated that the most significant impact on LHV are the coal and steam feed rates. As coal rate increases, the syngas production rate increases while the aeration and instrument purge nitrogen flow rates remain constant. Therefore, the nitrogen constituent of the syngas decreases (less dilution), and the syngas LHV increases. During oxygen-blown gasification, there is very little nitrogen that enters the gasifier with the oxidant, resulting in increases in the LHV.

Increasing the steam flow decreases the LHV by diluting the syngas with water. One way to combine the effects of changing in the mode of operation, as well as the change in steam and coal feed rates is to determine the overall percent of oxygen of all the gas that enters the Transport Gasifier. The overall percent O₂ is calculated by the following formula:

$$\text{Overall \%O}_2 = \frac{0.21 * \text{air} + \text{oxygen}}{\text{air} + \text{oxygen} + (\text{pure nitrogen}) + \text{steam}}$$

All flows are in moles per hour. At the PSDF, a large amount of pure nitrogen is fed to the gasifier for instrument purges, coal and sorbent transport, and equipment purges. During air-blown gasification about half of the nitrogen flowing through the gasifier comes from these pure nitrogen flows, and the remainder comes from the nitrogen in the air. When operating in oxygen-blown gasification, the nitrogen in the gasifier is predominantly due to the pure nitrogen flows.

The TC16 overall percentages of oxygen are listed in Table 3.1-3 and shown in Figure 3.1-7. The values ranged from 11.3 to 13.1 percent O₂ in air-blown mode with PRB coal and from 13.6 to 17.0 percent O₂ in oxygen-blown mode. During air-blown mode with lignite, the values ranged from 9.4 to 11.4 percent O₂ as a result of less oxygen in the fuel. The overall percent O₂ is higher in oxygen-blown gasification than in air-blown gasification due to less nitrogen dilution.

The TC16 LHV data are plotted against overall percent O₂ in feed on Figure 3.1-7. For comparison, the curves of previous PRB data (TC06, TC07, TC08, TC10, TC12, TC13, TC14, and TC15), previous Hiawatha bituminous data (TC09), previous Falkirk lignite (TC11), and previous Freedom Lignite (TC13) are included. The general trend for TC16 compares well with PRB data generated over the past several runs. The heating values were higher for PRB than for either of the lignites at the same overall percent O₂, but lower than the Hiawatha Bituminous.

Plotted on Figure 3.1-8 are the raw LHV data versus the air-to-coal ratio for the air-blown operating periods in TC16. The trend lines on the figure indicate the historical trends for previous PRB and Freedom Lignite testing. Note the decline in LHV as the air-to-coal ratio increases. The LHV values were slightly lower in TC16 than in previous test runs. All TC16 raw LHV data are plotted versus the oxygen-to-coal ratio in Figure 3.1-9.

Projected Syngas Heating Values. The PSDF Transport Gasifier produces syngas of a lower quality than a commercial sized gasifier due to:

- The use of recycle gas rather than nitrogen for aeration and PCD backpulsing.
- A lower heat loss per pound coal gasified because of a lower surface area to volume ratio.
- A smaller number of instruments and instrument purges.
- A cold gas cleanup train to remove contaminants and moisture from the syngas.

For details on the projected LHV calculation, see Appendix 5.

The projected LHV and adjusted syngas composition at the gas turbine inlet for each operating period are given in Table 3.1-5, along with the projected CO/CO₂ ratio. The projected LHV were between 117 and 140 Btu/SCF for air-blown operation and between 242 and 263 Btu/SCF for oxygen-blown operation using PRB coal. During air-blown mode with lignite coal, the projected LHV ranged from 103 to 118 Btu/SCF.

3.1.4 Gasifier Solids Analyses

During TC16, solid samples were taken from the following locations:

- Coal feed system (FD0210).
- Sorbent feed system (FD0220).
- Transport Gasifier standpipe (circulating solids).
- Transport Gasifier loop seal downcomer.
- Continuous fine ash depressurization system (CFAD or FD0540).
- PCD inlet (in situ).

The solids sample locations are shown in Figure 3.4-1 of the TC12 Topical Report. All solid samples were analyzed for chemical composition and particle size.

Table 3.1-6 gives the average proximate, ultimate and ash mineral analyses for PRB and Freedom lignite, as well as the average molar ratios for coal calcium to sulfur ratio (Ca/S) and the standard deviation for the samples analyzed as sampled from FD0210. Note the higher standard deviation in the lignite analyses than in the PRB analyses, as well as the higher sulfur and sodium content of the lignite when compared to the PRB. The coal carbon and moisture contents are shown in Figure 3.1-10. Both the moisture and the carbon content of the PRB coal remained essentially constant during TC16. The lignite carbon content steadily decreased and the lignite moisture increased for the last 100 hours of TC16.

Figure 3.1-11 shows the coal sulfur and ash as sampled from coal feeder FD0210 during TC16. The sulfur content remained between 0.22 and 0.31 weight percent for the PRB coal and from 0.34 and 1.52 weight percent for the Freedom lignite. The ash content for PRB coal was relatively constant, at around 5 weight percent, while the ash content for lignite varied from 6 to 11 weight percent.

The higher heating value (HHV) and LHV of the coal are given on Figure 3.1-12 with the TC16 average value given in Table 3.1-6. The coal HHV is determined using a bomb calorimeter. The calorimeter condenses all the coal combustion moisture as liquid water. The LHV is calculated by subtracting the heat of vaporization of the coal moisture from the HHV. Since heat recovery steam generators do not recover the coal syngas moisture heat of vaporization, the LHV is a more useful measure of coal heating value. The heating values for the PRB coal were fairly constant throughout TC16; however, the heating values for lignite continuously decreased.

FD0220 was used during TC16 to feed dolomite, limestone, or sand into the Transport Gasifier. The average analyses of the limestone and dolomite samples are given in Table 3.1-7.

The chemical compositions of the solid compounds produced by the Transport Gasifier were determined based on the chemical analysis and the following assumptions:

1. All carbon dioxide measured is from CaCO_3 , hence moles CO_2 measured = moles CaCO_3 .
2. All sulfide sulfur measured is from CaS .
3. All calcium not taken by CaS and CaCO_3 is from CaO .
4. All magnesium is from MgO .
5. Total carbon measured is the sum of organic and inorganic (CO_2) carbon. The organic carbon is the total carbon minus the inorganic carbon (CO_2).
6. Inerts are the sum of the BaO , P_2O_5 , FeO , K_2O , and TiO_2 concentrations.

Both elemental sulfur (ultimate analysis) and ash inert sulfur contents were measured. It is assumed that no FeS is formed in the Transport Gasifier and that all the sulfur in the standpipe and PCD fines solids is present as CaS . Thermodynamically, some FeS formation is possible, but most of the captured sulfur should be in the form of CaS due to the larger concentration of calcium than iron in the system.

Table 3.1-8 gives the TC16 standpipe solids analyses. These solids re-circulate through the mixing zone, riser, and standpipe. Typically, the properties of these solids change slowly with time. The standpipe composition data taken throughout TC16 vary due to changes in type and amounts of coal and sorbent fed.

Figure 3.1-13 shows the standpipe SiO_2 , CaO , and Al_2O_3 contents versus run time. As shown in the figure, sand was only added once after the run started, which was around 480 hours. The standpipe solids silica content was typically between 31 and 38 weight percent. Figure 3.1-13 demonstrates the replacement of startup sand by coal ash and sorbent. The standpipe started TC16 high in SiO_2 and low Al_2O_3 and CaO . As the run progressed, the SiO_2 decreased and the Al_2O_3 and CaO increased. Periodic sand addition returned the standpipe solids composition to near the start up level concentrations.

The standpipe solids organic carbon content was negligible for PRB and lignite throughout the entire run. The organic carbon is the total carbon in the solids minus inorganic carbon measured as CO_2 . Based on previous experience, the standpipe organic carbon content is a very inaccurate measurement because the value comes from a difference of two small values that are nearly equal.

The standpipe solids CaS content was negligible for all of the PRB standpipe samples. The calcium in the PRB coal ash captured a minimal amount of sulfur which is consistent with previous PRB test data. However, the CaS content for the lignite samples ranged from 0.11 to 1.66 percent as a result of the increased sulfur content in the lignite. The standpipe CaCO_3 averaged approximately 3 percent for PRB and 30 percent for lignite during TC16. The value was higher during the lignite testing because the lower temperature lignite testing did not calcine as much of the dolomite as the higher temperature PRB testing.

Table 3.1-9 lists the loop seal solids samples analyses. Figure 3.1-14 shows the CaO , SiO_2 , and Al_2O_3 contents of the loop seal solids samples. As shown, the SiO_2 content began to stabilize until sand was added at approximately 500 hours. The Al_2O_3 content also began to increase as the run progressed due to the increasing amount of inert in the coal. The CaO content remained relatively stable, around 12 percent, after 350 hours. The MgO , Fe_2O_3 , and other inert contents are not plotted, but they follow the same trend as the Al_2O_3 . The loop seal sodium oxide concentration was constant during the PRB testing and was between 0.7 and 2.5 percent Na_2O during lignite testing. The loop seal Na_2O increased from 2.1 to 5.3 percent Na_2O during the lignite testing.

Figure 3.1-15 shows the organic carbon (total carbon minus CO_2 carbon) contents for the loop seal solids. The carbon content of the solids is higher in the loop seal than the standpipe because a greater percentage of the smaller carbon particles are collected by the cyclone rather than the disengager.

The loop seal solids CaS content was negligible for all PRB samples and began to increase, as expected, for the lignite samples due to the increased sulfur content in the coal. The CaS content during the lignite period averaged 0.5 percent. The loop seal CaCO_3 was relatively constant for PRB and lignite averaging 6 and 20 percent, respectively.

Figure 3.1-16 plots the organic carbon for the PCD solids sampled from the new CFAD solids removal system (FD0540). Since FD0540 was in use during the run, none of the samples were taken from the FD0520 lock hopper solids removal system. The complete solids analysis and the organic carbon content for the PCD fines samples are given in Table 3.1-10. In situ PCD inlet particulate solid samples were also analyzed.

The in situ organic carbon contents are compared with the PCD solids samples on Figure 3.1-16. The in situ organic carbon solids were usually lower than the PCD solids samples organic carbon, but the trends still followed one another.

The carbon content remained between 20 and 50 percent throughout the test run, with the exception of the FD0540 sample taken at hour 51. This sample had a carbon content of 16 percent and was taken just after a period of sand addition.

Figure 3.1-17 and Table 3.1-10 show the amounts of SiO_2 and Al_2O_3 in the PCD solids as sampled from FD0540. The in situ solids concentrations for SiO_2 and Al_2O_3 are also plotted on the figure. Both the in situ and the actual measurements for SiO_2 and Al_2O_3 agreed well throughout the test campaign.

The SiO_2 PCD fines concentrations are a function of the efficiency of the disengager and cyclone as well as the SiO_2 concentration of circulating solids in the gasifier. The SiO_2 in the PCD fines comes from coal ash and sand. The SiO_2 PCD fines concentration slowly decreased after sand addition to relatively stable values, which were about 20 percent for PRB and 10 percent for lignite.

Since only a minimal amount of Al_2O_3 is in sand, the PCD fines Al_2O_3 content comes predominantly from the coal ash. The PCD fines Al_2O_3 concentration remained low, between 6 and 11 percent, throughout the test run.

Figure 3.1-18 shows the CaCO_3 and CaS concentrations of the PCD solids from FD0540 and in situ samples. The concentrations for CaO , CaS , and CaCO_3 are also listed in Table 3.1-10. Most of the in situ samples CaCO_3 and CaS concentrations agreed well with the PCD solids samples except for the CaCO_3 samples after hour 447. The PCD fines calcium concentration ranges were greater than the TC15 values due to the addition of limestone and dolomite.

The PCD fines calcination is defined as:

$$\% \text{Calcination} = \frac{\% \text{CaO}}{\% \text{CaO} + \% \text{CaCO}_3}$$

The PCD fines calcination data are plotted on Figure 3.1-19. The PCD fines calcination ranged from 28 to 87 percent for PRB and 5 to 34 percent for lignite. Previous PRB runs indicate calcination percentages between 60 and 80 percent, which was consistent with TC16. The cause of the lower calcination during the lignite operation was the lower operating temperature of the lignite testing.

The calcium sulfation is defined as:

$$\% \text{Sulfation} = \frac{\% \text{CaS}}{\% \text{CaO} + \% \text{CaCO}_3 + \% \text{CaS}}$$

The PCD fines sulfation is plotted on Figure 3.1-19 with the PCD fines calcination. The PCD fines sulfation varied between 0.8 and 10.6 percent for PRB and 1.9 and 8.8 percent for lignite throughout the test run. These values indicate that only a minimal amount of sulfur capture took place.

Solids Sample Comparison. The standpipe solids, the loop seal solids, and the PCD fines solids analysis comparison shows how the solids composition changes through the process.

Figure 3.1-20 compares the organic carbon content of the standpipe, loop seal, and PCD fines solids samples. The PCD solids carbon content ranged from 16 to 47 percent. The loop seal organic carbon content ranged between 0 and 5 percent, and the standpipe solids carbon content was between 0 and 2 percent. The data seem to indicate that the carbon is contained in small particles

which are only partially captured by the disengager. The cyclone separates a larger portion of the carbon, and the PCD collects the rest. As expected, most of the loop seal organic carbon data was between the standpipe and PCD fines data.

Figure 3.1-21 compares the calcium concentration between the standpipe, the loop seal, and the PCD solids samples. The calcium content was the generally highest for the standpipe solids after the first 200 hours, ranging from 3 to 25 percent. However, the loop seal calcium content values were equal to the standpipe calcium values for the last 300 hours, ranging between 12 and 24 percent. The PCD calcium content remained the lowest throughout TC16 (except for the first 200 hours), ranging from 6.7 to 19.2 percent. The standpipe and loop seal calcium contents were relatively constant after the first 200 hours of the test campaign, while the PCD calcium content changed, probably due to the dilution caused by the frequent limestone, dolomite, and sand additions. (Note that the calcium is distributed between the compounds CaO , CaCO_3 , and CaS .)

The silica entering the process primarily remains in the gasifier, since the sand particle size is greater than that of the standpipe solids. Figure 3.1-22 shows that the standpipe solids had the highest silica content, closely followed by the loop seal solids. The PCD solids had the lowest silica content, but it fluctuated, becoming higher during periods of poor solids collection performance and limestone, dolomite, or sand addition. As the gasifier operates sand eventually breaks down by attrition and exits through the cyclone before being collected by the PCD. Typically, a small amount of sand is lost through the cyclone, so spent solids slowly replace sand, providing sand addition is infrequent. As shown in Figure 3.1-22, two steady state times were achieved, one around 250 hours and another at approximately 700 hours.

Solids Particle Size. The TC16 Sauter mean diameter (SMD) and mass mean diameter (MMD) particle sizes of the coal sampled from coal feeder FD0210 are plotted on Figure 3.1-23. The PRB coal SMD particle size ranged from 119 to 297 microns for PRB and 128 to 304 microns for lignite. The average MMD for PRB coal was 274 microns, with a standard deviation of 68 microns, and the average MMD for lignite was 318 microns, with a standard deviation of 41 microns.

Figure 3.1-24 plots the SMD and MMD of the solids samples from sorbent feeder FD0220. The solids samples were limestone, dolomite, sand, or a mixture. The limestone sample at hour 51 was approximately 8 microns SMD and 17 microns MMD. There was a large variation between the SMD and MMD for the dolomite samples possibly due to mixture with sand. The vendor specifications for the sand were 150 microns for both the MMD and SMD.

Figure 3.1-25 plots both the coal feed percent above 1,180 microns (coarse particles) and percent below 45 microns (fines). A large amount of 1,180 micron particles increases the difference between the SMD and the MMD, because the SMD is a surface area average. Therefore, the larger particles with less surface area per pound have a weaker effect on the SMD than the MMD, where the larger particles skew the MMD due to their higher weight per particle. The average percent above 1,180 microns during TC16 for PRB was 5.7 percent with a standard deviation of 3.7 and the average was 5.2 percent with a standard deviation of 1.5 for lignite. The percent above 1,180 varied during the entire test run, between 1 and 21 percent for PRB and 3 and 8 percent for lignite. The high spikes in coarse solids correspond to the high MMD particle size.

In past testing, a high fines content in the feed coal resulted in an increased number of coal feeder outages due the packing of coal fines in the coal feed system lock vessel. These problems did not occur in TC16. The PRB average percent below 45 microns was 8.4 percent, with a standard deviation of 4.3, lower than typical PRB test runs. The lignite average percent below 45 microns was 5.7 percent, with a standard deviation of 3.6. The samples taken during the beginning of the test run had much more fines as compared to the middle of the test run. During the beginning of the test run, the values were consistently above 5 percent; however, even during these periods, the coal lock vessel did not experience any packing.

The TC16 standpipe solids particle sizes are given in Figure 3.1-26. The PRB standpipe solids particle sizes increased during the first 400 hours of the run as the system stabilized over time. The particle size did decrease at various times during the run due to sand and sorbent addition. During the first PRB oxygen-blown testing, the standpipe solids particle size slowly increased from the sand particle size (150 microns) to 280 microns (SMD) at about 300 hours. During the second oxygen-blown testing, after sand was added, the standpipe solids particle size increased quickly from 150 to 225 microns (SMD) in less than 100 hours. The standpipe solids particle size was steady for the first air-blown period and decreasing for the second air-blown period. During the lignite testing, the standpipe solids decreased for about 100 hours and then were steady at 135 microns (SMD) for the remaining 100 hours of TC16.

The percent of standpipe solids greater than 600 micron particles and the percent less than 45 microns are plotted on Figure 3.1-27. Figure 3.1-27 shows that the increase in particle size during the first oxygen-blown period was due to an increase in coarse particles and that the decrease in lignite standpipe particle size was due to an increase in fine particles. The coarse particles increased for the first 250 hours of TC16 and then were constant for the remainder to TC16. There was concern that the high sodium lignite standpipe particles would tend to agglomerate during TC16. When the standpipe solids particle size decreased and the amount of standpipe coarse particles also seemed to decrease, it appeared that operational changes made to prevent agglomeration were successful. The average fines content (less than 45 microns) was around 0.7 and 4.2 percent for PRB and lignite, respectively.

For some of the previous test campaigns the gasifier recirculating solids achieved a steady particle size, typically between 165 and 205 microns SMD as shown in Table 3.1-11. During TC16, the standpipe solids reached a steady state composition of 230 microns for PRB and 135 microns for lignite. By comparison, the startup sand is around 150 microns SMD. For tests that reached steady state the standpipe particle size slowly increased asymptotically to reach the steady state value. The TC16 maximum standpipe particle size was consistent with those of the previous PRB and lignite test campaigns as shown in Table 3.1-11.

The particle sizes of the loop seal solids are as shown in Figure 3.1-28. Both the SMD and the MMD of the loop seal solids varied widely. The SMD ranged from 7.1 to 147.7 microns, while the MMD varied from 16.5 to 200.7 microns.

Figure 3.1-29 plots the SMD and MMD for the PCD solids sampled from the FD0540 ash removal system as well as the situ solids recovered during the PCD inlet sampling. Overall, the trend for the in situ solids particle sizes agreed with the trend for the particle sizes of the solids collected from the

hopper. However, the absolute particle sizes did not always agree. Possible reasons for the disagreement are sand addition and particle segregation. In order to collect a representative sample of gasification ash, most of the in situ samples were collected during periods when sand was not being added to the gasifier. The hopper samples were often taken at times when sand was being added.

The PCD fines SMD started TC16 around 12 microns, and typically ranged from 5 to 16 microns for the remainder of the run, with one outlier at 19 microns. The MMD was typically around 10 microns larger than the SMD for most of the samples and followed the same trends as the SMD particle size. The in situ PCD inlet MMD solids particle size also showed the same trend of disagreement with the FD0540 solids MMD particle size during the spikes. Overall, the TC16 PCD fines particle size was consistent with the particle size of historical PRB and lignite test campaigns as shown in Table 3.1-11. The particle size was lower than it was in the two previous test campaigns, indicating that the cyclone performance had improved.

Particle Size Comparison. Figure 3.1-30 plots the solids SMD particle sizes, including the coal, standpipe solids, loop seal solids, and PCD fines from FD0540. The coal fed to the Transport Gasifier averaged 203 microns SMD for PRB, approximately the same size as the standpipe solids. For lignite, the average coal particle size was 241 microns, where the average standpipe particle size was only approximately 195 microns. The loop seal solids are typically more coarse than the PCD solids, but less coarse than the standpipe solids. Occasionally, the particle size of the PCD solids is greater than that of the loop seal solids. During the PRB operation, the loop seal particle size varied between the PCD fines and the standpipe particle size, while for the lignite operation, the loop seal particle size was the same as the standpipe particle size.

Standpipe and PCD Fines Bulk Densities. The standpipe, loop seal, PCD in situ, and FD0540 fines bulk densities are given in Figure 3.1-31. The PRB standpipe solids bulk density averaged approximately 86 pounds per cubic foot, a value close to that of pure sand. The lignite standpipe solids bulk density decreased from 85 to 64 pounds per cubic foot. Falling between the standpipe and PCD bulk density data, the data for the loop seal solids averaged 68 pounds per cubic foot, but varied considerably more than the standpipe solids data. The PCD solids had the lowest average values, around 14.5 pounds per cubic foot for PRB and 30 pounds per cubic foot for lignite based on PCD solids samples from FD0540. All of the in situ PCD data points were in agreement with the FD0540 sample data.

The minimum standpipe solids densities for past PSDF gasification test campaigns are shown in Table 3.1-11. Also listed are the average SMD particle size and standard deviation of the PCD fines for all previous gasification test campaigns.

3.1.5 Carbon Conversion

Carbon conversion is defined as the percent of fuel carbon that is gasified to CO, CO₂, CH₄, C₂H₆, and higher hydrocarbons versus the amount of carbon that is rejected by the gasifier with the PCD and gasifier solids. In a typical flow sheet for integrated gasification combine cycle (IGCC), the unconverted carbon from the gasifier or PCD is burned in a combustor or sent to a landfill for disposal.

The carbon conversion can be calculated several different ways, the most accurate method divides the carbon content in the syngas by the total carbon exiting the gasifier (from both solid and gas streams). Table 3.1-12 gives the carbon conversions for the 55 test periods, while Figure 3.1-32 shows the carbon conversion versus time. The carbon conversion ranged from 92 to 96 percent for PRB and from 83 to 87 percent for low temperature lignite during the test run. During TC16, the mode of operation (air or oxygen) did not appear to affect the carbon conversion. In the past, the carbon conversion for air blown has been slightly higher than the carbon conversion for oxygen blown due to the lack of oxygen nozzles in the UMZ. These nozzles were installed and testing in TC16, allowing for higher riser temperatures, and resulting in a higher carbon conversion.

Since the gasifier operated over a wide range of temperatures in TC16, the effects on carbon conversion were notable. Figure 3.1-33 illustrates the relationship between the mixing zone temperature and the carbon conversion for TC16 for both PRB and Freedom Lignite in both air- and oxygen-blown modes. Also shown in the figure are the data for TC13 as a comparison. The PRB data had the highest carbon conversion for TC16 and did not appear to vary based on air- or oxygen-blown mode of operation. Since the gasifier operated at a lower temperature using lignite in TC16 than it did using PRB, the resulting carbon conversion was lower.

As shown in Figure 3.3-33, the highest carbon conversions were those of the TC13 lignite at higher temperatures. Unfortunately, due to ash agglomeration, operations above 1,500°F were impractical for extended periods of time. Thus, a majority of the TC13 lignite periods were at very low temperatures with correspondingly low carbon conversions. During TC16, the use of dolomite allowed higher temperatures in the gasifier without agglomeration, resulting in an increase in carbon conversion of around 5 to 10 percent.

The average carbon conversions of Powder River Basin, Hiawatha seam bituminous from Utah, and Falkirk and Freedom lignites from North Dakota are compared on Figure 3.1-34 for both air and oxygen operation. These data came from Test Runs TC06 through TC13 and TC15 through TC16. The graph does not include TC14 data, since the poor performance of the solids collection systems caused abnormally low carbon conversion. The low temperature Freedom lignite carbon conversion data are plotted separately from the high temperature Freedom lignite carbon conversion data to illustrate that significantly lower temperatures adversely affect the carbon conversion. Air-blown operation yielded a slightly higher carbon conversion than oxygen-blown operation for all fuels. Only for Hiawatha bituminous coal, however, is the difference between air- and oxygen-blown modes larger than the standard deviation of the average values.

Falkirk lignite had the highest average carbon conversion of the four coals tested. PRB and Freedom lignite had about the same average carbon conversion, while Hiawatha Bituminous had the lowest average carbon conversion. Although the data in Figure 3.1-34 show general trends in carbon conversion over test runs, the values obtained are the result of operating over a small range of conditions for all fuels except PRB and lignite fuels.

3.1.6 Gasification Efficiencies

Gasification efficiency is defined as the percentage of the entering energy that is converted to potentially useful syngas energy. Two types of gasification efficiencies have been defined: the cold gasification efficiency and the hot gasification efficiency. The cold gasification efficiency is the amount of energy fed that is available to a gas turbine as syngas latent heat. The hot gasification efficiency is the percentage of total energy fed that is available to produce electricity. The total energy to produce electricity includes the syngas latent heat recovered in a gas turbine plus the sensible heat recovered in a steam turbine.

The cold gasification efficiency is plotted in Figure 3.1-35 and is listed in Table 3.1-12. During TC16, the cold gasification efficiencies were as high as 61.3 percent for PRB and 43.9 percent for low temperature lignite. During the air-blown operating periods, the cold gas efficiencies ranged from 47.3 to 55.2 percent for PRB and from 30.9 to 42.2 percent for lignite. The oxygen-blown periods had cold gas efficiencies between 51.8 and 61.3 percent for PRB. Since less nitrogen had to be heated to gasifier temperature in oxygen-blown mode, the efficiencies tend to be higher.

The hot gasification efficiency assumes that the sensible heat of the syngas can be recovered in a heat recovery steam generator, so the hot gasification efficiency is always higher than the cold gasification efficiency. The hot gasification efficiency is the latent and the sensible heat of the syngas exiting the gasifier divided by the total amount of energy entering the gasifier, including the latent heat of the coal and the sensible heats of the air, and steam. The hot gasification efficiency is plotted in Figure 3.1-36 and shown in Table 3.1-12. The efficiency boundary for the values found in Figure 3.1-36 and Table 3.1-12 is the gasifier system itself, not including any downstream equipment.

The air blown hot gas efficiencies were between 80.0 and 85.8 percent for PRB and from 67.1 to 73.6 percent for low temperature lignite. The oxygen blown hot gasification efficiencies ranged from 82.0 to 86.6 percent.

The two main sources of efficiency losses are the gasifier heat loss and the latent heat of the PCD solids. The gasifier heat loss of 3.5 million Btu/hr was about 10 percent of the feed energy, while the total energy of the PCD solids was from 3 to 15 percent of the feed energy (the higher numbers occurring during the periods of low coal feed rate).

A commercial gasifier will be more efficient than the PSDF gasifier due to the use of recycle gas and lower heat losses. The heat loss as a percentage of energy fed will be much smaller in a commercially sized gasifier. While the Transport Gasifier does not recover the latent heat of the PCD solids, this latent heat could be recovered in a combustor. The total enthalpy of the PCD solids can be decreased by decreasing both the PCD solids carbon content (heating value) and the PCD solids rate (by improving solids collection efficiency).

Gasification efficiencies can be calculated from the commercially projected gas heating values and adjusted flow rates that were determined when calculating the projected heating value. The commercially projected cold gasification efficiencies are listed on Table 3.1-12 for all of the operating periods.

The projected efficiencies are calculated assuming an adiabatic gasifier, since zero heat loss was one of the assumptions in determining the corrected LHV in Section 3.1.3. The projected cold gasification efficiencies were from 68.2 to 72.4 percent for PRB and from 50.0 to 58.8 percent for lignite in air-blown mode. The projected cold gasification efficiencies were between 74.5 and 79.1 percent in oxygen-blown mode. The use of recycle gas and the lower heat losses in the commercial projection are the factors that increase efficiency.

Table 3.1-1 Operating Periods (Page 1 of 2)

Operating Period	Start Time	End Time	Duration Hours	Operating Period	
				Average Time	Relative Hours
TC16-1	7/15/04 10:15	7/15/04 19:30	9:15	7/15/04 14:52	20
TC16-2	7/15/04 23:00	7/16/04 7:15	8:15	7/16/04 3:07	32
TC16-3	7/16/04 17:00	7/16/04 21:15	4:15	7/16/04 19:07	48
TC16-4	7/17/04 12:30	7/17/04 16:15	3:45	7/17/04 14:22	67
TC16-5	7/17/04 21:30	7/18/04 1:00	3:30	7/17/04 23:15	76
TC16-6	7/18/04 2:15	7/18/04 5:30	3:15	7/18/04 3:52	81
TC16-7	7/18/04 22:00	7/19/04 2:15	4:15	7/19/04 0:07	101
TC16-8	7/19/04 17:00	7/19/04 19:45	2:45	7/19/04 18:22	119
TC16-9	7/20/04 2:00	7/20/04 12:00	10:00	7/20/04 7:00	132
TC16-10	7/21/04 20:00	7/22/04 1:30	5:30	7/21/04 22:45	172
TC16-11	7/22/04 14:00	7/22/04 17:00	3:00	7/22/04 15:30	189
TC16-12	7/22/04 17:15	7/22/04 23:30	6:15	7/22/04 20:22	193
TC16-13	7/23/04 1:15	7/23/04 6:45	5:30	7/23/04 4:00	201
TC16-14	7/23/04 18:00	7/24/04 2:45	8:45	7/23/04 22:22	219
TC16-15	7/24/04 12:00	7/24/04 17:00	5:00	7/24/04 14:30	236
TC16-16	7/25/04 8:45	7/25/04 12:30	3:45	7/25/04 10:37	256
TC16-17	7/25/04 13:00	7/25/04 16:30	3:30	7/25/04 14:45	260
TC16-18	7/25/04 17:00	7/26/04 2:45	9:45	7/25/04 21:52	267
TC16-19	7/26/04 8:30	7/26/04 12:15	3:45	7/26/04 10:22	279
TC16-20	7/27/04 16:00	7/27/04 23:45	7:45	7/27/04 19:52	313
TC16-21	7/27/04 23:45	7/28/04 7:15	7:30	7/28/04 3:30	321
TC16-22	7/28/04 23:00	7/29/04 9:15	10:15	7/29/04 4:07	345
TC16-23	7/29/04 19:30	7/30/04 6:00	10:30	7/29/04 23:45	366
TC16-24	7/30/04 23:00	7/31/04 2:00	3:00	7/31/04 0:30	390
TC16-25	7/31/04 3:00	7/31/04 9:45	6:45	7/31/04 6:22	395
TC16-26	7/31/04 16:15	7/31/04 20:45	4:30	7/31/04 18:30	408
TC16-27	7/31/04 22:00	8/1/04 2:30	4:30	8/1/04 0:15	413
TC16-28	8/1/04 4:00	8/1/04 10:00	6:00	8/1/04 7:00	420
TC16-29	8/1/04 10:30	8/1/04 14:30	4:00	8/1/04 12:30	425
TC16-30	8/1/04 15:30	8/1/04 18:30	3:00	8/1/04 17:00	430

Notes:

1. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
2. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.

Table 3.1-1 Operating Periods (Page 2 of 2)

Operating Period	Start Time	End Time	Duration Hours	Operating Period	
				Average Time	Relative Hours
TC16-31	8/1/04 21:00	8/2/04 2:45	5:45	8/1/04 23:52	437
TC16-32	8/2/04 9:15	8/2/04 18:00	8:45	8/2/04 13:37	451
TC16-33	8/3/04 7:00	8/3/04 13:30	6:30	8/3/04 10:15	471
TC16-34	8/10/04 14:00	8/10/04 18:00	4:00	8/10/04 16:00	487
TC16-35	8/10/04 22:30	8/11/04 1:30	3:00	8/11/04 0:00	495
TC16-36	8/11/04 2:00	8/11/04 6:45	4:45	8/11/04 4:22	499
TC16-37	8/11/04 16:15	8/12/04 4:15	12:00	8/11/04 22:15	517
TC16-38	8/12/04 17:00	8/12/04 20:30	3:30	8/12/04 18:45	538
TC16-39	8/13/04 12:00	8/13/04 15:00	3:00	8/13/04 13:30	557
TC16-40	8/15/04 6:45	8/15/04 14:00	7:15	8/15/04 10:22	601
TC16-41	8/16/04 4:00	8/16/04 11:45	7:45	8/16/04 7:52	623
TC16-42	8/16/04 19:00	8/17/04 6:45	11:45	8/17/04 0:52	640
TC16-43	8/17/04 13:00	8/17/04 16:00	3:00	8/17/04 14:30	654
TC16-44	8/17/04 18:30	8/18/04 2:00	7:30	8/17/04 22:15	661
TC16-45	8/18/04 6:15	8/19/04 8:15	2:00	8/18/04 19:15	682
TC16-46	8/19/04 12:00	8/19/04 19:30	7:30	8/19/04 15:45	703
TC16-47	8/19/04 22:00	8/20/04 7:00	9:00	8/20/04 2:30	714
TC16-48	8/20/04 23:45	8/21/04 7:30	7:45	8/21/04 3:37	739
TC16-49	8/21/04 16:30	8/22/04 7:30	15:00	8/22/04 0:00	759
TC16-50	8/22/04 14:30	8/23/04 2:45	12:15	8/22/04 20:37	780
TC16-51	8/23/04 4:00	8/23/04 16:00	12:00	8/23/04 10:00	793
TC16-52	8/23/04 21:30	8/24/04 3:30	6:00	8/24/04 0:30	808
TC16-53	8/24/04 6:30	8/24/04 9:15	2:45	8/24/04 7:52	815
TC16-54	8/24/04 13:00	8/24/04 19:45	6:45	8/24/04 16:22	823
TC16-55	8/24/04 20:00	8/24/04 22:45	2:45	8/24/04 21:22	828

Notes:

1. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
2. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.

Table 3.1-2 Operating Conditions (Page 1 of 2)

Operating Periods	Average Relative Hours	Riser Exit Temperature Zone °F	Pressure PI287 psig	Coal Rate ² lb/hr	Sorbent Rate ⁴ lb/hr	Air Rate lb/hr	Oxygen Rate lb/hr	Syngas Rate lb/hr	Steam Rate ³ lb/hr	Nitrogen Rate lb/hr	PCD Inlet Temperature TI458 °F	PCD Solids Rate lb/hr
TC16-1	20	1,712	217	3,700	0	12,000	0	20,600	800	5,800	750	360
TC16-2	32	1,703	217	3,600	0	11,900	0	20,900	900	6,200	739	380
TC16-3	48	1,669	145	3,000	0	1,300	2,100	14,400	3,500	5,500	677	390
TC16-4	67	1,675	145	3,100	0	1,100	2,100	13,900	3,100	5,700	666	400
TC16-5	76	1,664	145	3,100	0	1,200	2,100	14,500	3,600	5,500	672	410
TC16-6	81	1,658	145	3,200	0	1,200	2,100	14,400	3,400	5,600	669	410
TC16-7	101	1,676	145	3,300	0	1,100	2,100	14,500	3,300	5,500	682	420
TC16-8	119	1,653	155	3,400	110	1,300	2,200	14,100	3,000	5,400	665	450
TC16-9	132	1,651	155	3,400	140	1,000	2,100	14,300	3,100	5,500	700	470
TC16-10	172	1,649	155	3,600	150	1,300	2,200	14,500	3,000	5,500	680	390
TC16-11	189	1,678	155	3,400	440	1,900	2,200	15,400	3,200	5,600	703	380
TC16-12	193	1,672	155	3,400	200	1,300	2,300	15,000	3,200	5,700	699	350
TC16-13	201	1,679	155	3,000	120	1,100	2,100	14,400	3,100	5,900	711	310
TC16-14	219	1,675	155	3,700	150	1,300	2,300	15,100	3,200	5,700	723	280
TC16-15	236	1,671	155	3,500	80	1,300	2,300	14,700	3,100	5,600	730	280
TC16-16	256	1,669	155	3,700	70	1,300	2,400	15,100	3,000	5,800	697	290
TC16-17	260	1,677	155	3,800	120	1,400	2,400	15,600	3,100	5,800	704	290
TC16-18	267	1,663	155	3,900	100	1,000	2,300	15,900	3,100	6,100	699	290
TC16-19	279	1,665	155	3,900	100	1,100	2,400	15,500	3,100	5,700	696	290
TC16-20	313	1,664	155	3,700	100	1,200	2,300	16,000	3,300	6,100	697	400
TC16-21	321	1,663	155	3,600	110	1,000	2,200	16,000	3,300	6,200	695	430
TC16-22	345	1,676	155	3,300	100	1,000	2,100	16,300	3,400	6,500	700	270
TC16-23	366	1,674	155	3,200	110	1,300	2,100	15,900	3,400	6,500	701	220
TC16-24	390	1,671	217	3,700	0	11,600	0	21,900	1,500	6,400	745	220
TC16-25	395	1,674	217	3,700	0	11,400	0	21,700	1,500	6,400	747	230
TC16-26	408	1,682	217	3,700	0	11,300	0	20,500	600	6,300	734	240
TC16-27	413	1,683	217	3,400	0	10,700	0	19,600	500	6,300	728	250
TC16-28	420	1,679	217	3,400	0	10,700	0	19,700	600	6,300	726	250
TC16-29	425	1,679	217	3,100	0	10,100	0	18,800	600	6,200	720	260
TC16-30	430	1,675	217	3,100	0	10,000	0	18,600	500	6,200	717	260

Notes:

1. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
2. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.
3. Steam rate calculated by hydrogen balance.
4. TC16-8 to TC16-13 fed limestone; TC16-14 to TC16-23, TC16-34 to TC16-38, and TC16-40 to TC16-53 fed dolomite.

Table 3.1-2 Operating Conditions (Page 2 of 2)

Operating Periods	Average Relative Hours	Riser Exit Temperature Zone °F	Pressure PI287 psig	Coal Rate ² lb/hr	Sorbent Rate ⁴ lb/hr	Air Rate lb/hr	Oxygen Rate lb/hr	Syngas Rate lb/hr	Steam Rate ³ lb/hr	Nitrogen Rate lb/hr	PCD Inlet Temperature TI458 °F	PCD Solids Rate lb/hr
TC16-31	437	1,674	217	2,700	0	8,900	0	17,700	600	6,500	712	270
TC16-32	451	1,700	217	4,000	0	12,000	0	21,800	800	6,200	750	280
TC16-33	471	1,686	217	3,400	0	10,900	0	20,000	900	6,400	734	280
TC16-34	487	1,648	154	3,000	60	2,500	1,800	15,800	3,200	5,900	713	290
TC16-35	495	1,648	162	3,600	70	1,500	2,300	16,300	3,600	6,000	733	290
TC16-36	499	1,657	162	3,700	80	1,400	2,300	15,800	3,200	6,000	740	290
TC16-37	517	1,683	191	4,100	70	1,500	2,600	17,000	3,400	6,300	788	270
TC16-38	538	1,683	191	4,000	100	1,400	2,600	17,600	3,800	6,400	795	290
TC16-39	557	1,670	186	4,400	0	900	2,100	17,600	4,000	6,100	789	360
TC16-40	601	1,698	220	3,800	210	12,600	0	22,600	1,100	6,300	779	340
TC16-41	623	1,700	226	4,100	140	13,400	0	23,600	1,200	6,200	772	330
TC16-42	640	1,688	226	4,500	150	13,600	0	24,200	1,200	6,200	768	330
TC16-43	654	1,440	150	2,900	200	7,900	0	17,300	1,500	5,700	669	620
TC16-44	661	1,444	150	2,900	290	8,300	0	18,000	1,600	5,900	668	610
TC16-45	682	1,462	150	2,800	260	8,200	0	17,600	1,600	5,800	667	660
TC16-46	703	1,434	150	3,100	350	7,900	0	17,700	1,700	6,000	661	730
TC16-47	714	1,449	150	2,400	340	7,500	0	17,000	1,800	6,200	659	640
TC16-48	739	1,459	150	3,400	340	9,000	0	18,800	1,600	6,000	668	610
TC16-49	759	1,475	152	3,500	320	9,000	0	18,400	1,200	6,000	666	640
TC16-50	780	1,472	152	3,500	300	9,200	0	18,300	1,200	6,000	665	670
TC16-51	793	1,465	152	3,500	330	9,300	0	18,400	1,300	5,900	667	680
TC16-52	808	1,463	152	3,400	300	9,100	0	18,100	1,200	6,000	665	650
TC16-53	815	1,481	152	3,500	310	9,300	0	18,300	1,200	5,900	674	640
TC16-54	823	1,472	159	4,000	330	9,900	0	19,500	1,300	6,000	682	630
TC16-55	828	1,460	159	4,000	330	9,600	0	19,100	1,200	6,100	678	630

Notes:

1. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
2. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.
3. Steam rate calculated by hydrogen balance.
4. TC16-8 to TC16-13 fed limestone; TC16-14 to TC16-23, TC16-34 to TC16-38, and TC16-40 to TC16-53 fed dolomite.

Table 3.1-3 Raw Gas Composition, Molecular Weight, and Heating Value (Page 1 of 2)

Operating Period	Average Relative Hour	H ₂ O ¹ Mole %	CO Mole %	H ₂ Mole %	CO ₂ Mole %	CH ₄ Mole %	C ₂ H ₆ Mole %	Argon Mole %	N ₂ Mole %	Total Mole %	Wet Syngas MW lb/Mole	Overall %O ₂ %	Syngas CO/CO ₂ Ratio	Wet Syngas LHV Btu/SCF
TC16-1	20	9.1	9.9	7.8	8.3	1.2	0.0	0.6	63.1	100.0	26.3	13.1	1.2	65
TC16-2	32	9.6	9.2	7.6	8.4	1.2	0.0	0.5	63.5	100.0	26.3	12.6	1.1	61
TC16-3	48	29.7	7.2	13.2	12.5	1.5	0.0	0.1	35.8	100.0	23.4	14.8	0.6	73
TC16-4	67	28.2	7.8	13.3	12.6	1.5	0.0	0.1	36.4	100.0	23.6	15.2	0.6	76
TC16-5	76	30.8	7.0	13.1	12.6	1.5	0.0	0.1	35.0	100.0	23.4	14.9	0.6	72
TC16-6	81	29.5	7.4	13.3	12.7	1.6	0.0	0.1	35.4	100.0	23.5	15.1	0.6	75
TC16-7	101	28.5	8.1	13.6	12.8	1.7	0.0	0.1	35.1	100.0	23.5	15.5	0.6	80
TC16-8	119	26.6	8.9	14.1	13.2	2.0	0.0	0.1	35.2	100.0	23.6	16.3	0.7	85
TC16-9	132	26.5	8.9	14.0	13.2	2.1	0.0	0.1	35.2	100.0	23.6	15.9	0.7	86
TC16-10	172	27.8	7.8	13.3	13.0	1.7	0.0	0.1	36.3	100.0	23.7	16.3	0.6	78
TC16-11	189	26.9	8.1	12.9	12.8	1.8	0.0	0.1	37.4	100.0	23.8	16.0	0.6	78
TC16-12	193	27.4	8.0	13.0	12.9	1.9	0.0	0.1	36.7	100.0	23.7	16.2	0.6	79
TC16-13	201	28.6	6.6	11.7	12.2	1.5	0.0	0.1	39.4	100.0	23.9	14.9	0.5	67
TC16-14	219	27.4	8.2	13.3	12.8	1.9	0.0	0.1	36.2	100.0	23.6	16.6	0.6	80
TC16-15	236	27.5	8.2	13.4	12.8	1.8	0.0	0.1	36.2	100.0	23.6	16.8	0.6	79
TC16-16	256	27.0	8.3	13.3	12.9	1.9	0.0	0.1	36.5	100.0	23.7	17.0	0.6	80
TC16-17	260	26.4	8.5	13.3	12.8	1.8	0.0	0.1	37.1	100.0	23.7	16.7	0.7	81
TC16-18	267	26.0	8.4	13.3	12.7	1.8	0.0	0.1	37.6	100.0	23.8	16.2	0.7	80
TC16-19	279	26.6	8.6	13.6	12.9	1.7	0.0	0.1	36.4	100.0	23.7	16.9	0.7	81
TC16-20	313	27.8	7.4	12.7	12.5	1.5	0.0	0.1	37.8	100.0	23.7	15.5	0.6	73
TC16-21	321	27.8	7.3	12.6	12.4	1.5	0.0	0.1	38.4	100.0	23.8	15.0	0.6	72
TC16-22	345	28.1	6.5	11.5	12.0	1.4	0.0	0.1	40.5	100.0	24.0	14.1	0.5	65
TC16-23	366	28.7	6.1	11.2	11.9	1.3	0.0	0.2	40.8	100.0	24.0	14.2	0.5	62
TC16-24	390	13.6	6.3	7.2	9.2	1.1	0.0	0.5	62.0	100.0	26.2	11.8	0.7	51
TC16-25	395	13.9	6.0	6.9	9.2	1.2	0.0	0.5	62.3	100.0	26.2	11.8	0.7	49
TC16-26	408	9.0	8.4	6.8	8.4	1.3	0.0	0.6	65.6	100.0	26.6	12.6	1.0	58
TC16-27	413	9.0	8.0	6.5	8.2	1.2	0.0	0.5	66.6	100.0	26.7	12.4	1.0	54
TC16-28	420	9.1	8.1	6.6	8.5	1.1	0.0	0.5	66.1	100.0	26.7	12.4	1.0	55
TC16-29	425	9.2	7.4	6.2	8.3	1.1	0.0	0.6	67.1	100.0	26.7	12.2	0.9	52
TC16-30	430	9.1	7.4	6.2	8.3	1.1	0.0	0.6	67.4	100.0	26.7	12.2	0.9	51

Notes:

1. The H₂O concentration was estimated using a correlation between the in-situ samples and gasifier temperature.
2. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
3. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.

Table 3.1-3 Raw Gas Compositions, Molecular Weight, and Heating Value (Page 2 of 2)

Operating Period	Average Relative Hour	H ₂ O ¹ Mole %	CO Mole %	H ₂ Mole %	CO ₂ Mole %	CH ₄ Mole %	C ₂ H ₆ Mole %	Argon Mole %	N ₂ Mole %	Total Mole %	Wet Syngas MW lb/Mole	Overall %O ₂ %	Syngas CO/CO ₂ Ratio	Wet Syngas LHV Btu/SCF
TC16-31	437	8.8	7.0	5.8	8.1	1.0	0.0	0.5	68.7	100.0	26.9	11.3	0.9	48
TC16-32	451	10.1	8.6	7.5	8.6	1.2	0.0	0.6	63.4	100.0	26.4	12.7	1.0	60
TC16-33	471	10.8	7.3	6.9	8.6	1.1	0.0	0.5	64.7	100.0	26.5	12.1	0.9	52
TC16-34	487	27.6	5.5	10.2	11.9	1.4	0.0	0.2	43.2	100.0	24.4	14.2	0.5	59
TC16-35	495	28.8	6.9	12.3	12.8	1.9	0.0	0.1	37.3	100.0	23.8	15.3	0.5	73
TC16-36	499	26.7	7.6	12.4	12.8	2.0	0.0	0.1	38.4	100.0	23.9	15.9	0.6	77
TC16-37	517	25.7	9.1	13.5	12.9	2.2	0.0	0.1	36.5	100.0	23.8	16.9	0.7	86
TC16-38	538	27.9	8.0	13.0	12.8	1.9	0.0	0.1	36.3	100.0	23.7	16.0	0.6	79
TC16-39	557	28.3	8.8	14.3	13.4	2.1	0.0	0.1	33.0	100.0	23.4	13.6	0.7	87
TC16-40	601	11.0	7.8	7.1	8.9	1.2	0.0	0.5	63.5	100.0	26.4	12.7	0.9	55
TC16-41	623	10.7	8.5	7.5	8.9	1.2	0.0	0.5	62.7	100.0	26.3	13.0	1.0	60
TC16-42	640	10.9	8.4	7.6	9.1	1.5	0.0	0.5	62.0	100.0	26.3	13.1	0.9	61
TC16-43	654	17.4	3.0	6.1	9.8	0.9	0.0	0.5	62.3	100.0	26.2	10.2	0.3	35
TC16-44	661	17.3	3.0	6.1	9.8	0.9	0.0	0.5	62.4	100.0	26.2	10.3	0.3	35
TC16-45	682	18.1	2.6	5.4	9.7	0.7	0.0	0.5	62.9	100.0	26.3	10.3	0.3	30
TC16-46	703	19.9	2.3	5.5	9.7	0.8	0.0	0.5	61.3	100.0	26.1	9.9	0.2	30
TC16-47	714	20.3	1.9	4.7	9.3	0.6	0.0	0.4	62.7	100.0	26.2	9.4	0.2	25
TC16-48	739	18.2	2.9	6.0	9.8	0.8	0.0	0.5	61.7	100.0	26.2	10.6	0.3	34
TC16-49	759	16.2	3.4	6.0	9.8	0.8	0.0	0.5	63.2	100.0	26.3	11.0	0.3	35
TC16-50	780	16.5	3.4	6.1	9.9	0.8	0.0	0.5	62.9	100.0	26.3	11.1	0.3	35
TC16-51	793	16.4	3.5	6.3	9.9	0.9	0.0	0.5	62.5	100.0	26.3	11.2	0.4	37
TC16-52	808	15.7	3.6	6.3	9.8	0.9	0.0	0.5	63.2	100.0	26.3	11.1	0.4	37
TC16-53	815	15.8	3.7	6.2	9.8	0.8	0.0	0.5	63.2	100.0	26.3	11.3	0.4	37
TC16-54	823	16.4	3.9	6.8	10.1	0.9	0.0	0.5	61.4	100.0	26.2	11.4	0.4	40
TC16-55	828	16.1	3.8	6.7	10.0	0.9	0.0	0.5	62.0	100.0	26.2	11.3	0.4	40

Notes:

1. The H₂O concentration was estimated using a correlation between the in-situ samples and gasifier temperature.
2. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
3. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.

Table 3.1-4 Syngas Sulfur Concentration (Page 1 of 2)

Operating Period	Average Relative Hour	Wet AI419J H ₂ S ppm	Syngas Total Reduced Sulfur ³ ppm	Sorbent Rate ⁴ lb/hr	Sulfur Capture %
TC16-1	20	82	348	0	23
TC16-2	32	96	312	0	22
TC16-3	48	316	443	0	8
TC16-4	67	299	393	0	6
TC16-5	76	351	421	0	6
TC16-6	81	386	453	0	6
TC16-7	101	432	457	0	6
TC16-8	119	436	465	110	6
TC16-9	132	435	448	140	7
TC16-10	172	386	319	150	6
TC16-11	189	391	364	440	8
TC16-12	193	410	341	200	8
TC16-13	201	364	292	120	10
TC16-14	219	420	359	150	7
TC16-15	236	414	359	80	7
TC16-16	256	421	378	70	6
TC16-17	260	410	349	120	6
TC16-18	267	422	338	100	7
TC16-19	279	417	336	100	8
TC16-20	313	385	308	100	10
TC16-21	321	374	294	110	11
TC16-22	345	387	306	100	7
TC16-23	366	410	354	110	5
TC16-24	390	325	280	0	10
TC16-25	395	313	255	0	10
TC16-26	408	304	286	0	13
TC16-27	413	305	259	0	18
TC16-28	420	249	274	0	20
TC16-29	425	288	286	0	22
TC16-30	430	286	285	0	23

Notes:

1. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
2. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.
3. Syngas total reduced sulfur (TRS) calculated from syngas combustor SO₂ analyzer data.
4. TC16-8 to TC16-13 fed limestone; TC16-14 to TC16-23, TC16-34 to TC16-38, and TC16-40 to TC16-53 fed dolomite.

Table 3.1-4 Syngas Sulfur Concentration (Page 2 of 2)

Operating Period	Average Relative Hour	Wet AI419J H ₂ S ppm	Syngas Total Reduced Sulfur ³ ppm	Sorbent Rate ⁴ lb/hr	Sulfur Capture %
TC16-31	437	284	241	0	27
TC16-32	451	305	202	0	26
TC16-33	471	293	250	0	19
TC16-34	487	186	341	60	18
TC16-35	495	324	356	70	16
TC16-36	499	342	348	80	15
TC16-37	517	348	341	70	7
TC16-38	538	347	299	100	5
TC16-39	557	410	379	0	5
TC16-40	601	254	238	210	21
TC16-41	623	254	197	140	24
TC16-42	640	250	212	150	20
TC16-43	654	770	938	200	12
TC16-44	661	861	962	290	11
TC16-45	682	833	893	260	11
TC16-46	703	883	943	350	13
TC16-47	714	782	860	340	12
TC16-48	739	835	806	340	10
TC16-49	759	735	785	320	16
TC16-50	780	751	810	300	21
TC16-51	793	764	752	330	23
TC16-52	808	733	769	300	21
TC16-53	815	686	733	310	24
TC16-54	823	773	802	330	23
TC16-55	828	785	756	330	25

Notes:

1. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
2. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.
3. Syngas total reduced sulfur (TRS) calculated from syngas combustor SO₂ analyzer data.
4. TC16-8 to TC16-13 fed limestone; TC16-14 to TC16-23, TC16-34 to TC16-38, and TC16-40 to TC16-53 fed dolomite.

Table 3.1-5 Projected¹ Gas Compositions, Molecular Weight, and Heating Value (Page 1 of 2)

Operating Period	Average Relative Hour	H ₂ O Mole %	CO Mole %	H ₂ Mole %	CO ₂ Mole %	CH ₄ Mole %	C ₂ H ₆ Mole %	Argon Mole %	N ₂ Mole %	Total Mole %	Syngas MW lb/mole	Overall % O ₂	Syngas CO/CO ₂ Ratio	Syngas LHV Btu/SCF
TC16-1	20	1.0	18.3	16.8	10.7	2.5	0.0	0.6	50.2	100.0	25	17.0	1.7	128
TC16-2	32	1.0	17.6	17.3	11.2	2.5	0.0	0.6	49.9	100.0	25	16.5	1.6	127
TC16-3	48	1.0	26.4	42.8	24.7	5.1	0.0	0.0	0.0	100.0	20	19.4	1.1	250
TC16-4	67	1.0	27.8	42.4	23.8	5.0	0.0	0.0	0.0	100.0	20	22.2	1.2	252
TC16-5	76	1.0	25.5	42.9	25.6	5.0	0.0	0.0	0.0	100.0	20	19.4	1.0	246
TC16-6	81	1.0	26.4	42.4	25.0	5.3	0.0	0.0	0.0	100.0	20	21.0	1.1	250
TC16-7	101	1.0	27.7	41.6	24.1	5.6	0.0	0.0	0.0	100.0	20	21.4	1.1	255
TC16-8	119	1.0	28.4	40.9	23.9	6.0	0.0	0.0	0.0	100.0	20	23.2	1.2	258
TC16-9	132	1.0	29.2	40.1	23.3	6.4	0.0	0.0	0.0	100.0	20	23.0	1.3	263
TC16-10	172	1.0	26.7	41.4	25.4	5.6	0.0	0.0	0.0	100.0	21	24.5	1.0	251
TC16-11	189	1.0	27.1	40.5	25.7	5.7	0.0	0.0	0.0	100.0	21	21.8	1.1	251
TC16-12	193	1.0	27.3	40.6	25.1	6.0	0.0	0.0	0.0	100.0	21	22.9	1.1	255
TC16-13	201	1.0	26.7	41.8	25.0	5.5	0.0	0.0	0.0	100.0	20	21.7	1.1	251
TC16-14	219	1.0	27.3	41.0	24.7	6.0	0.0	0.0	0.0	100.0	21	23.5	1.1	256
TC16-15	236	1.0	27.1	41.4	24.9	5.6	0.0	0.0	0.0	100.0	20	22.6	1.1	252
TC16-16	256	1.0	27.3	40.8	25.1	5.8	0.0	0.0	0.0	100.0	21	24.2	1.1	254
TC16-17	260	1.0	28.0	40.6	24.7	5.7	0.0	0.0	0.0	100.0	21	24.1	1.1	254
TC16-18	267	1.0	28.7	40.4	24.1	5.8	0.0	0.0	0.0	100.0	21	25.6	1.2	257
TC16-19	279	1.0	28.1	40.9	24.6	5.4	0.0	0.0	0.0	100.0	21	26.2	1.1	252
TC16-20	313	1.0	26.8	41.6	25.4	5.2	0.0	0.0	0.0	100.0	21	24.4	1.1	248
TC16-21	321	1.0	27.6	41.4	24.8	5.2	0.0	0.0	0.0	100.0	21	23.9	1.1	251
TC16-22	345	1.0	27.4	41.4	25.0	5.2	0.0	0.0	0.0	100.0	21	22.8	1.1	250
TC16-23	366	1.0	25.8	42.1	26.0	5.1	0.0	0.0	0.0	100.0	21	22.3	1.0	245
TC16-24	390	1.0	13.4	18.5	14.5	2.7	0.0	0.6	49.5	100.0	25	15.1	0.9	118
TC16-25	395	1.0	12.9	18.1	14.7	2.8	0.0	0.6	49.9	100.0	25	15.2	0.9	117
TC16-26	408	1.0	17.6	16.8	11.5	3.0	0.0	0.6	49.6	100.0	25	17.4	1.5	130
TC16-27	413	1.0	17.4	17.2	11.4	2.8	0.0	0.6	49.7	100.0	25	17.4	1.5	129
TC16-28	420	1.0	17.6	16.9	11.8	2.7	0.0	0.6	49.5	100.0	25	17.5	1.5	128
TC16-29	425	1.0	17.0	17.2	11.8	2.9	0.0	0.6	49.5	100.0	25	17.1	1.4	129
TC16-30	430	1.0	17.0	17.4	11.6	2.7	0.0	0.6	49.6	100.0	25	17.3	1.5	128

Notes:

- Adjustments are based on the following assumptions: that only air nitrogen is in the syngas, the gasifier is adiabatic, and syngas is at the turbine inlet after the syngas cleanup processes.
- TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
- TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.

Table 3.1-5 Projected¹ Gas Compositions, Molecular Weight, and Heating Value (Page 2 of 2)

Operating Period	Average Relative Hour	H ₂ O Mole %	CO Mole %	H ₂ Mole %	CO ₂ Mole %	CH ₄ Mole %	C ₂ H ₆ Mole %	Argon Mole %	N ₂ Mole %	Total Mole %	Syngas MW lb/mole	Overall %O ₂ %	Syngas CO/CO ₂ Ratio	Syngas LHV Btu/SCF
TC16-31	437	1.0	18.6	19.2	12.0	3.0	0.0	0.6	45.8	100.0	25	16.5	1.5	140
TC16-32	451	1.0	17.1	17.4	12.0	2.6	0.0	0.6	49.3	100.0	25	17.0	1.4	127
TC16-33	471	1.0	15.5	17.6	12.5	2.5	0.0	0.6	50.3	100.0	25	16.2	1.2	121
TC16-34	487	1.0	24.0	40.5	28.7	5.8	0.0	0.0	0.0	100.0	21	21.7	0.8	242
TC16-35	495	1.0	25.7	40.1	26.8	6.5	0.0	0.0	0.0	100.0	21	22.5	1.0	252
TC16-36	499	1.0	27.5	39.5	25.3	6.7	0.0	0.0	0.0	100.0	21	25.2	1.1	259
TC16-37	517	1.0	30.0	38.7	23.8	6.6	0.0	0.0	0.0	100.0	21	26.3	1.3	263
TC16-38	538	1.0	28.7	39.3	25.0	6.1	0.0	0.0	0.0	100.0	21	25.0	1.2	256
TC16-39	557	1.0	29.9	38.8	24.0	6.3	0.0	0.0	0.0	100.0	21	27.4	1.3	261
TC16-40	601	1.0	15.4	16.5	12.7	2.6	0.0	0.6	51.2	100.0	25	16.3	1.2	119
TC16-41	623	1.0	16.0	16.3	12.4	2.5	0.0	0.6	51.2	100.0	25	16.4	1.3	120
TC16-42	640	1.0	15.7	16.2	12.8	2.9	0.0	0.6	50.8	100.0	25	16.6	1.2	122
TC16-43	654	1.0	8.2	22.3	20.2	3.2	0.1	0.5	44.6	100.0	25	13.0	0.4	118
TC16-44	661	1.0	8.2	21.6	19.9	2.9	0.1	0.6	45.9	100.0	25	12.9	0.4	113
TC16-45	682	1.0	7.3	20.4	20.2	2.5	0.1	0.6	48.0	100.0	26	13.0	0.4	103
TC16-46	703	1.0	6.7	21.6	21.5	2.7	0.1	0.6	45.8	100.0	25	13.1	0.3	107
TC16-47	714	1.0	6.0	21.5	22.0	2.6	0.1	0.6	46.3	100.0	26	11.4	0.3	103
TC16-48	739	1.0	7.6	19.9	19.5	2.5	0.1	0.6	48.9	100.0	26	13.9	0.4	103
TC16-49	759	1.0	8.8	19.8	18.7	2.4	0.1	0.6	48.7	100.0	26	14.9	0.5	106
TC16-50	780	1.0	8.4	19.2	18.3	2.4	0.1	0.6	50.1	100.0	26	14.8	0.5	103
TC16-51	793	1.0	8.5	19.3	18.2	2.4	0.1	0.6	49.9	100.0	26	14.5	0.5	104
TC16-52	808	1.0	9.1	19.9	18.1	2.5	0.1	0.6	48.8	100.0	25	14.4	0.5	108
TC16-53	815	1.0	9.1	19.3	17.8	2.4	0.1	0.6	49.8	100.0	26	15.0	0.5	105
TC16-54	823	1.0	9.2	19.6	18.0	2.5	0.1	0.6	49.1	100.0	25	15.0	0.5	107
TC16-55	828	1.0	9.2	19.9	18.0	2.6	0.1	0.6	48.7	100.0	25	15.2	0.5	109

Notes:

- Adjustments are based on the following assumptions: that only air nitrogen is in the syngas, the gasifier is adiabatic, and syngas is at the turbine inlet after the syngas cleanup processes.
- TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
- TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were lignite.

Table 3.1-6 Coal Analysis

	Powder River Basin		Freedom Lignite	
	Average Value ¹	Standard Deviation	Average Value ¹	Standard Deviation
Moisture, wt%	21.39	1.50	22.57	2.39
Carbon, wt%	55.67	0.92	51.07	3.27
Hydrogen ² , wt%	3.63	0.15	3.35	0.26
Nitrogen, wt%	0.74	0.02	0.71	0.02
Oxygen, wt%	13.41	0.83	13.95	1.54
Sulfur, wt%	0.26	0.03	0.71	0.36
Ash, wt%	5.02	0.55	8.35	1.99
Volatiles, wt%	33.51	2.33	31.33	2.77
Fixed Carbon, wt%	40.21	2.15	38.47	5.61
Higher Heating Value, Btu/lb	9,249	140	8,394	578
Lower Heating Value, Btu/lb	8,925	135	8,084	556
CaO, wt %	1.24	0.08	1.15	0.78
SiO ₂ , wt %	1.57	0.25	1.54	1.12
Al ₂ O ₃ , wt %	0.82	0.09	0.89	0.62
MgO, wt %	0.31	0.02	0.40	0.27
Na, wt % in ash	1.26	0.35	4.92	1.89
Na ₂ O, wt% in ash	1.70	0.48	6.63	2.55
Fe ₂ O ₃ , wt% in ash	6.23	0.37	9.49	1.88
Ca/S, mole/mole	2.79	0.40	0.69	0.53

Notes:

1. All analyses are as sampled at FD0210.
2. Hydrogen in coal is reported separately from hydrogen in moisture.

Table 3.1-7 Sorbent Analysis

Compound	Dolomite		Limestone ³	
	Weight %	Standard Deviation	Weight %	Standard Deviation
CaCO ₃ , wt %	52.98	1.06	74.80	2.05
MgCO ₃ , wt %	42.72	1.30	17.36	1.89
CaSO ₄ , wt%	0.24	0.06	0.61	0.05
SiO ₂ , wt %	2.72	0.87	6.46	3.02
Al ₂ O ₃ , wt %	0.58	0.17	0.44	0.09
Other inerts ²	0.05	0.01	0.24	0.09
H ₂ O, wt %	0.37	0.09	0.09	0.05
Total	99.7		100.0	

Notes:

1. All samples taken from FD0220 sorbent feeder.
2. Other inerts consist of P₂O₅, Na₂O, K₂O, Fe₂O₃, and TiO₂.
3. Limestone values are typical values taken from previous test campaigns.

Table 3.1-8 Standpipe Solids Analysis

Sample Number	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	Na ₂ O Wt. %	Other Inerts ¹ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon ² Wt. %	Total Wt. %
AB15137	7/15/2004 18:00	23	87.3	4.0	0.4	3.2	2.4	1.2	1.9	0.8	0.0	101.2
AB15167	7/16/2004 18:00	47	86.5	4.2	0.6	3.2	2.4	0.0	3.2	1.0	0.0	101.0
AB15170	7/17/2004 18:00	71	87.2	4.0	0.4	3.4	2.6	0.0	2.3	0.8	0.0	100.8
AB15172	7/18/2004 10:00	87	86.9	4.0	0.6	3.4	2.8	0.0	2.1	0.7	0.0	100.5
AB15235	7/20/2004 18:00	143	74.6	5.8	0.9	3.9	3.0	0.0	10.7	2.0	0.0	101.0
AB15257	7/21/2004 10:00	159	68.8	6.1	1.0	4.2	2.6	0.0	15.6	2.4	0.0	100.7
AB15259	7/22/2004 2:00	175	61.0	6.8	1.2	4.8	2.5	0.0	21.1	3.3	0.0	100.6
AB15282	7/23/2004 2:00	199	48.4	8.1	1.4	4.8	3.0	0.0	30.6	4.5	0.0	101.0
AB15316	7/24/2004 2:00	223	44.4	9.2	1.5	5.0	2.7	0.0	32.2	5.4	0.0	100.4
AB15317	7/24/2004 10:00	231	42.3	9.8	1.7	6.2	2.3	0.0	32.7	5.5	0.0	100.4
AB15320	7/25/2004 10:00	255	40.0	10.8	1.9	6.3	3.2	0.0	32.3	6.4	1.9	102.7
AB15323	7/26/2004 10:00	279	37.1	11.5	1.8	6.6	3.2	0.0	32.1	7.9	0.0	100.2
AB15365	7/27/2004 10:00	303	35.5	11.2	2.2	6.4	3.1	0.0	33.1	9.2	0.0	100.7
AB15382	7/28/2004 2:00	319	34.9	11.0	2.0	6.6	2.4	0.0	33.7	9.3	0.0	99.9
AB15407	7/29/2004 10:00	351	34.4	11.6	1.6	6.5	2.6	0.0	33.6	9.8	0.0	100.1
AB15445	7/30/2004 10:00	375	33.2	11.8	1.7	7.0	2.5	0.0	33.8	10.2	0.0	100.3
AB15449	7/31/2004 18:00	407	33.3	11.7	1.2	6.7	2.4	0.0	34.1	10.0	0.0	99.5
AB15452	8/1/2004 18:00	431	34.0	13.4	1.6	7.6	2.9	0.0	31.4	9.8	0.0	100.7
AB15497	8/2/2004 18:00	455	34.3	12.6	1.8	7.1	2.8	0.0	31.8	8.7	0.7	99.7
AB13243	8/3/2004 10:00	471	34.0	12.7	1.8	7.3	3.3	0.0	32.1	9.2	0.0	100.4
AB13282	8/10/2004 10:00	481	58.8	8.6	1.0	5.6	2.9	0.0	18.1	5.7	0.0	100.6
AB15552	8/10/2004 18:00	489	63.9	7.9	1.0	5.0	3.9	0.0	14.3	4.3	0.5	100.8
AB15553	8/11/2004 2:00	497	51.7	9.8	1.2	5.8	3.5	0.0	22.0	6.6	0.0	100.8
AB15593	8/12/2004 18:00	537	43.2	11.1	1.9	6.6	3.2	0.0	26.1	8.4	0.5	100.9
AB15604	8/13/2004 13:00	556	38.8	12.4	1.8	6.7	3.4	0.0	27.7	9.7	0.0	100.6
AB15633	8/14/2004 10:00	577	36.8	12.6	2.1	6.8	3.0	0.0	29.1	10.3	0.0	100.8
AB15636	8/15/2004 10:00	601	35.9	11.8	2.4	6.8	3.7	0.0	29.9	10.2	0.0	100.7
AB15638	8/16/2004 2:00	617	34.3	11.2	2.0	6.8	3.5	0.0	31.6	11.1	0.0	100.6
AB15669	8/17/2004 2:00	641	33.6	11.9	2.1	6.5	2.6	0.0	32.5	11.4	0.3	100.9
AB15684	8/17/2004 10:00	649	32.3	11.0	1.8	6.3	3.3	0.0	32.7	11.1	0.4	98.8
AB15692	8/18/2004 2:00	665	35.9	11.1	2.5	6.0	3.5	0.0	29.8	10.2	0.8	99.8
AB15710	8/19/2004 2:00	689	29.8	10.4	3.5	6.0	17.3	0.0	22.6	12.4	0.0	102.0
AB15752	8/20/2004 2:00	713	27.5	9.4	3.9	5.7	21.8	0.1	19.2	13.2	0.2	100.9
AB15796	8/21/2004 2:00	737	23.5	8.0	3.8	5.0	29.6	0.0	14.7	15.1	0.0	99.7
AB15799	8/22/2004 2:00	761	22.8	8.3	4.4	5.2	37.3	0.1	10.2	15.4	0.0	103.6
AB15802	8/23/2004 2:00	785	24.5	8.5	5.4	5.4	31.9	0.7	11.6	15.4	0.0	103.4
AB15832	8/24/2004 2:00	809	22.6	7.8	5.0	4.6	39.7	0.5	7.7	16.6	0.0	104.4
AB15847	8/24/2004 22:30	830	25.3	8.8	5.9	5.5	33.1	1.7	8.7	15.5	0.0	104.4

Notes:

1. Other inerts consist of P₂O₅, FeO, K₂O, and TiO₂.
2. Organic carbon was below the detection limit when 0% was reported.
3. Samples AB15137 to AB15669 were sampled during PRB operation; Samples AB15684 to AB15847 were sampled during lignite operation.

Table 3.1-9 Loop Seal Solids Analysis

Sample Number	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	Na ₂ O Wt. %	Other Inerts ¹ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon Wt. %	Total Wt. %
AB15234	7/21/2004 2:00	151	67.4	6.3	0.7	2.0	16.8	0.0	7.7	3.0	0.0	103.9
AB15426	7/30/2004 10:00	375	34.2	12.5	1.6	1.9	13.1	0.0	24.1	11.4	0.2	98.9
AB15500	8/2/2004 18:00	455	39.2	14.1	1.9	1.9	15.5	0.2	17.1	7.0	4.0	100.9
AB15502	8/3/2004 10:00	471	39.2	13.8	2.1	2.0	5.2	0.1	23.5	7.5	3.7	97.1
AB15561	8/11/2004 10:00	505	60.5	9.3	1.3	1.9	6.2	0.0	13.1	5.0	5.1	102.4
AB15613	8/13/2004 18:00	561	37.4	12.3	1.7	1.9	9.8	0.0	22.4	11.3	3.1	99.8
AB15615	8/14/2004 10:00	577	35.8	12.1	1.9	1.8	8.1	0.1	24.1	11.7	3.1	98.7
AB15620	8/16/2004 2:00	617	34.8	11.6	1.7	1.7	5.6	0.1	28.0	12.5	1.7	97.7
AB15675	8/17/2004 2:00	641	32.5	11.8	1.6	1.6	2.5	0.1	32.2	12.6	1.1	96.0
AB15685	8/17/2004 10:00	649	31.0	11.7	2.1	1.4	7.6	0.4	26.3	12.2	5.0	97.7
AB15686	8/17/2004 18:00	657	29.0	10.9	2.3	1.3	14.2	0.2	23.0	13.9	4.1	98.9
AB15687	8/18/2004 2:00	665	28.1	10.5	2.9	1.2	15.0	0.3	22.1	14.2	0.0	94.3
AB15718	8/19/2004 2:00	689	26.7	9.8	4.0	1.1	17.7	0.1	21.3	15.1	0.6	96.3
AB15749	8/20/2004 2:00	713	23.8	8.6	3.8	0.9	23.6	0.1	18.6	17.3	2.2	98.9
AB15783	8/21/2004 2:00	737	22.1	8.1	3.5	0.8	26.5	0.1	17.5	17.9	1.7	98.2
AB15786	8/22/2004 2:00	761	23.4	8.4	4.4	0.8	23.7	0.5	18.2	17.3	1.3	97.9
AB15789	8/23/2004 2:00	785	24.6	8.7	4.8	0.8	18.0	0.8	19.4	17.4	1.7	96.2
AB15833	8/23/2004 18:00	801	24.7	8.7	5.2	0.8	25.6	0.5	14.7	16.5	1.3	98.0
AB15846	8/24/2004 22:30	830	25.2	9.4	5.3	0.8	21.8	2.1	14.9	16.5	1.5	97.5

Notes:

1. Other inerts consist of P₂O₅, FeO, K₂O, and TiO₂.
2. Samples AB15234 to AB15675 were sampled during PRB operation; Samples AB15685 to AB15846 were sampled during lignite operation.

Table 3.1-10 PCD Solids from FD0540 Analysis

Sample Number	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	Na ₂ O Wt. %	Other Inerts ¹ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic C (C-CO ₂) Wt. %	Total Wt. %	HHV Btu/lb	LHV Btu/lb
AB15131	7/15/2004 18:00	23	29.2	8.7	1.2	5.7	13.2	1.6	2.9	2.6	34.5	101.4	5,450	5,420
AB15134	7/16/2004 6:00	35	38.1	8.7	1.1	5.6	7.1	1.3	6.2	2.7	29.9	102.6	4,582	4,550
AB15178	7/16/2004 22:00	51	55.2	7.8	1.0	5.4	6.9	0.2	5.6	2.3	16.4	102.5	2,612	2,593
AB15182	7/17/2004 14:00	67	50.7	8.2	0.9	5.3	9.4	0.3	4.0	2.2	21.5	103.8	3,157	3,130
AB15186	7/18/2004 6:00	83	43.5	8.0	1.0	5.3	9.2	0.3	4.9	2.4	30.3	106.5	4,089	4,052
AB15213	7/19/2004 18:00	119	38.1	7.7	0.8	5.1	9.2	0.3	8.0	3.2	28.5	102.8	4,499	4,461
AB15225	7/20/2004 10:00	135	27.9	7.1	1.2	4.8	7.2	0.3	15.8	5.1	29.3	100.3	4,541	4,505
AB15250	7/22/2004 2:00	175	25.8	6.7	0.9	4.1	7.7	0.2	18.4	5.1	29.2	99.6	4,555	4,514
AB15276	7/22/2004 22:00	195	23.6	7.5	1.1	4.7	9.2	0.4	21.3	6.5	24.7	100.8	3,774	3,739
AB15287	7/23/2004 10:00	207	26.8	9.4	1.2	5.5	5.1	0.4	18.5	4.7	27.8	101.4	4,144	4,105
AB15302	7/24/2004 18:00	239	18.0	8.1	1.1	4.8	8.9	0.4	12.7	7.0	39.4	101.9	5,706	5,653
AB15308	7/25/2004 18:00	263	17.0	8.2	1.1	4.6	8.8	0.4	13.1	7.9	40.0	102.6	5,807	5,750
AB15345	7/26/2004 14:00	283	15.8	7.8	1.2	4.3	8.6	0.5	12.1	7.6	41.5	100.5	6,231	6,170
AB15376	7/27/2004 22:00	315	16.0	8.3	1.3	4.6	9.1	0.4	14.4	9.2	34.8	99.4	5,254	5,211
AB15378	7/28/2004 6:00	323	16.1	8.0	1.3	4.6	9.1	0.4	14.2	9.0	38.4	102.4	5,403	5,364
AB15404	7/29/2004 10:00	351	18.5	9.0	1.5	5.4	9.3	0.4	17.9	10.6	28.0	102.2	3,957	3,926
AB15421	7/30/2004 6:00	371	19.2	9.2	1.6	5.5	9.0	0.4	16.2	9.2	30.8	102.6	4,323	4,277
AB15467	7/31/2004 2:00	391	17.7	9.1	1.6	5.2	6.8	0.9	9.7	3.9	47.1	103.3	6,661	6,599
AB15469	7/31/2004 10:00	399	21.5	11.3	1.6	6.5	6.3	0.7	14.0	5.2	33.9	103.1	4,842	4,800
AB15475	8/1/2004 10:00	423	19.6	10.1	1.8	5.9	7.1	1.6	10.4	3.8	40.1	102.1	5,715	5,662
AB15481	8/2/2004 10:00	447	21.6	11.4	1.7	6.9	9.0	1.6	11.4	4.5	34.4	104.9	4,943	4,902
AB15507	8/3/2004 10:00	471	20.9	10.5	2.1	5.8	8.6	1.1	8.9	3.4	40.2	102.7	5,907	5,863
AB15556	8/10/2004 22:00	493	31.3	9.8	1.3	6.2	7.4	1.2	9.2	3.6	31.2	103.0	4,608	4,562
AB15578	8/12/2004 6:00	525	21.4	7.4	1.2	4.6	8.2	0.3	12.6	6.6	39.4	103.0	5,704	5,644
AB15587	8/12/2004 22:00	541	23.5	8.5	1.5	5.2	9.3	0.3	16.0	9.3	28.4	103.4	5,342	5,314
AB15612	8/13/2004 18:00	561	19.1	8.8	1.5	4.9	10.5	0.3	12.0	7.8	35.9	101.9	5,231	5,186
AB15649	8/15/2004 18:00	609	14.8	8.2	1.6	4.9	10.4	1.3	13.8	10.0	33.9	100.1	5,871	5,819
AB15676	8/16/2004 18:00	633	13.7	7.4	1.6	4.4	11.7	1.1	11.6	9.1	39.2	100.7	5,569	5,506
AB15679	8/17/2004 6:00	645	15.4	8.1	1.2	4.7	10.3	1.8	11.2	8.8	37.6	100.6	5,689	5,633
AB15693	8/17/2004 18:00	657	11.1	6.6	4.8	5.1	23.3	1.0	4.8	10.4	33.8	99.6	5,023	4,959
AB15695	8/18/2004 2:00	665	11.4	6.7	4.7	5.0	21.1	1.0	6.1	10.3	33.4	98.5	5,087	5,024
AB15721	8/18/2004 22:00	685	13.5	7.4	5.8	5.5	24.3	0.7	6.6	11.9	25.5	99.4	3,904	3,847
AB15755	8/19/2004 22:00	709	10.5	6.0	5.5	4.6	23.9	1.0	5.8	11.8	30.6	97.4	4,404	4,339
AB15758	8/20/2004 10:00	721	10.9	6.2	5.9	5.0	28.9	0.6	7.3	15.1	21.0	98.4	3,076	3,033
AB15806	8/21/2004 6:00	741	11.0	6.4	5.7	4.6	27.0	0.8	4.4	11.9	28.7	97.9	4,376	4,329
AB15812	8/22/2004 6:00	765	10.8	6.5	5.8	4.7	24.2	1.3	5.3	12.1	30.3	98.5	4,590	4,542
AB15817	8/23/2004 2:00	785	11.6	6.7	5.6	5.0	21.8	1.7	4.9	10.9	32.9	98.8	5,005	4,946
AB15841	8/23/2004 22:00	805	10.0	6.0	5.8	4.0	26.6	1.5	0.9	10.1	34.9	96.6	5,494	5,427
AB15843	8/24/2004 14:00	821	10.4	6.3	5.6	4.9	25.0	2.1	2.5	10.7	34.1	99.4	5,269	5,208

Notes:

1. Other inerts consist of P₂O₅, FeO, K₂O, and TiO₂.

2. Samples AB15131 to AB15679 were sampled during PRB operation; Samples AB15693 to AB15843 were sampled during lignite operation.

Table 3.1-11 Historical Standpipe and PCD Fines

Test Campaign	Fuel	Standpipe			PCD Fines			
		Maximum	Steady		Average	St. Dev.		
		Particle Size	State	Minimum Bulk	Particle Size	Particle Size	Average Bulk	St. Dev. Bulk
		SMD	SMD	Density	SMD	SMD	Density	Density
		microns	microns	lb/ft ³	microns	microns	lb/ft ³	lb/ft ³
TC06	Powder River Basin	204	165	80	10.8	1.1	24	4
TC07	Powder River Basin	191	175	80	10.2	1.1	28	8
TC07	Alabama Bituminous	232	none	66	16.2	3.2	32	7
TC08	Powder River Basin	250	205	77	13.1	3.2	25	7
TC09	Hiawatha Bituminous	233	180	76	15.7	4.6	29	12
TC10	Powder River Basin	280	none	76	10.7	3.6	23	7
TC11	Falkirk Lignite	200	200	75	12.3	2.4	36	3
TC12	Powder River Basin	300	none	76	9.8	2	18	6
TC13	Powder River Basin	165	165	81	10.4	1.4	18	4
TC13	Freedom Lignite Low Sodium	230	none	56	15.3	3.9	26	6
TC13	Freedom Lignite High Sodium, High Temp.	425	none	46	30.0	32.3	39	14
TC13	Freedom Lignite High Sodium, Low Temp.	457	none	67	13.9	2.3	26	5
TC14	Powder River Basin	220	none	84	18.7	14.6	27	14
TC15	Powder River Basin	156	none	79	10.7	1.3	20	4
TC16	Powder River Basin	288	280	75	11.4	2.7	17	3
TC16	Freedom Lignite, Low Temp.	173	135	64	11.0	1.8	32	4

Table 3.1-12 Carbon Conversion and Gasification Efficiencies (Page 1 of 2)

Operating Period1	Average Relative Hours	Carbon Conversion %	Efficiency		
			Raw		Projected ³
			Cold %	Hot %	Cold %
TC16-1	20	93.6	54.9	83.8	70.4
TC16-2	32	94.1	54.1	84.0	70.4
TC16-3	48	95.5	56.6	83.9	77.0
TC16-4	67	94.6	56.8	83.3	77.1
TC16-5	76	93.5	55.6	82.9	75.4
TC16-6	81	93.0	56.4	82.8	75.9
TC16-7	101	93.1	57.4	83.1	76.0
TC16-8	119	93.1	58.7	82.7	76.4
TC16-9	132	92.8	58.9	82.7	76.7
TC16-10	172	93.6	57.1	82.8	75.8
TC16-11	189	94.7	57.9	84.3	76.5
TC16-12	193	95.1	58.3	84.5	77.4
TC16-13	201	94.6	54.1	82.7	76.4
TC16-14	219	95.1	59.0	85.0	77.7
TC16-15	236	94.1	58.0	83.9	76.7
TC16-16	256	93.9	58.5	84.1	76.8
TC16-17	260	94.1	58.8	84.6	77.0
TC16-18	267	94.1	58.9	84.6	77.6
TC16-19	279	93.8	58.7	84.2	76.7
TC16-20	313	92.5	55.8	82.7	74.7
TC16-21	321	91.5	55.1	82.0	74.5
TC16-22	345	95.1	54.7	84.3	77.3
TC16-23	366	95.9	53.9	84.4	77.2
TC16-24	390	95.9	51.2	85.3	69.4
TC16-25	395	95.5	50.2	84.8	68.6
TC16-26	408	95.1	53.6	84.7	72.4
TC16-27	413	94.3	51.6	83.3	71.5
TC16-28	420	94.0	51.6	83.0	71.1
TC16-29	425	93.3	49.6	81.7	70.3
TC16-30	430	93.2	49.1	81.4	70.1

Notes:

1. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
2. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were Lignite.
3. Projection assumes that all nitrogen in the syngas is from air and that the gasifier is adiabatic.

Table 3.1-12 Carbon Conversion and Gasification Efficiencies (Page 2 of 2)

Operating Period ¹	Average Relative Hours	Carbon Conversion %	Efficiency		
			Raw		Projected ³
			Cold %	Hot %	Cold %
TC16-31	437	92.7	47.3	80.0	70.7
TC16-32	451	94.8	54.3	85.3	71.3
TC16-33	471	93.2	50.2	82.4	68.6
TC16-34	487	94.3	51.8	82.3	74.7
TC16-35	495	95.1	57.5	85.0	77.2
TC16-36	499	95.0	58.4	84.9	78.1
TC16-37	517	95.3	61.3	86.6	79.1
TC16-38	538	95.7	59.2	86.1	77.9
TC16-39	557	94.7	61.2	86.3	78.3
TC16-40	601	94.0	51.9	83.7	68.2
TC16-41	623	94.2	54.2	85.2	69.4
TC16-42	640	94.3	55.2	85.8	69.9
TC16-43	654	83.5	38.3	69.9	56.5
TC16-44	661	84.7	38.9	71.1	57.1
TC16-45	682	85.2	35.6	69.9	54.4
TC16-46	703	83.1	35.0	68.8	53.0
TC16-47	714	84.3	30.9	67.1	50.0
TC16-48	739	87.3	38.8	72.8	57.0
TC16-49	759	85.9	39.1	72.0	57.4
TC16-50	780	84.4	38.5	70.8	55.7
TC16-51	793	83.9	39.2	70.6	55.8
TC16-52	808	83.9	39.3	70.4	56.4
TC16-53	815	84.6	39.3	71.1	56.5
TC16-54	823	86.1	42.2	73.6	58.7
TC16-55	828	85.7	41.9	73.0	58.8

Notes:

1. TC16-1 to TC16-2, TC16-24 to TC16-33, and TC16-40 to TC16-55 were air blown; TC16-3 to TC16-23 and TC16-34 to TC16-39 were oxygen blown.
2. TC16-1 to TC16-42 were Powder River Basin coal; TC16-43 to TC16-55 were Lignite.
3. Projection assumes that all nitrogen in the syngas is from air and that the gasifier is adiabatic.

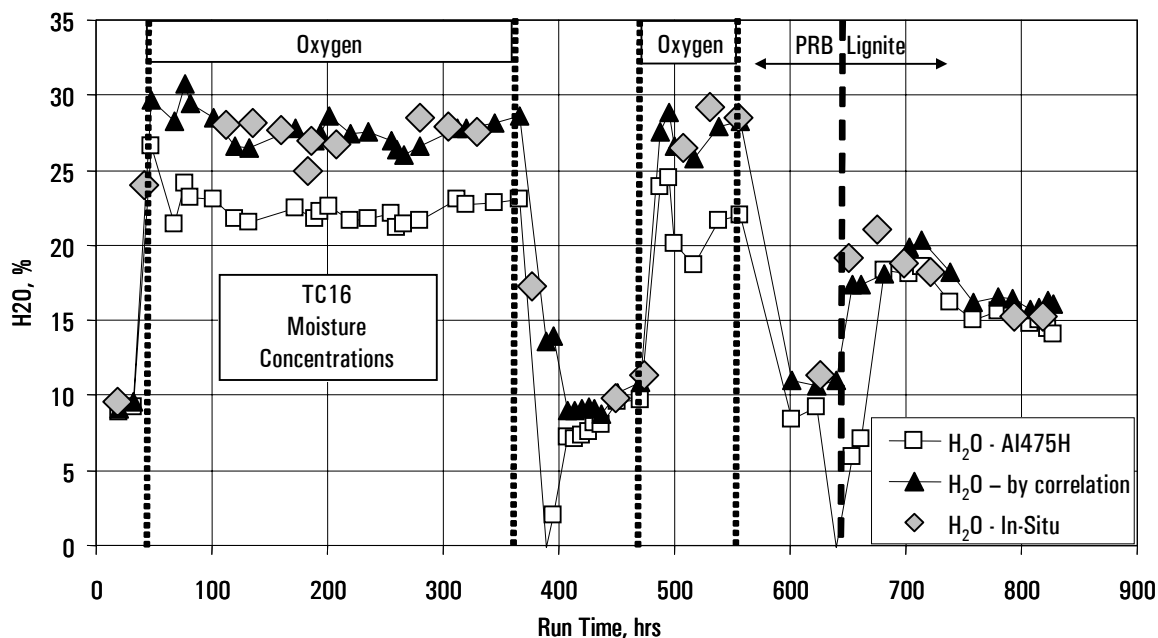


Figure 3.1-1 H₂O Data

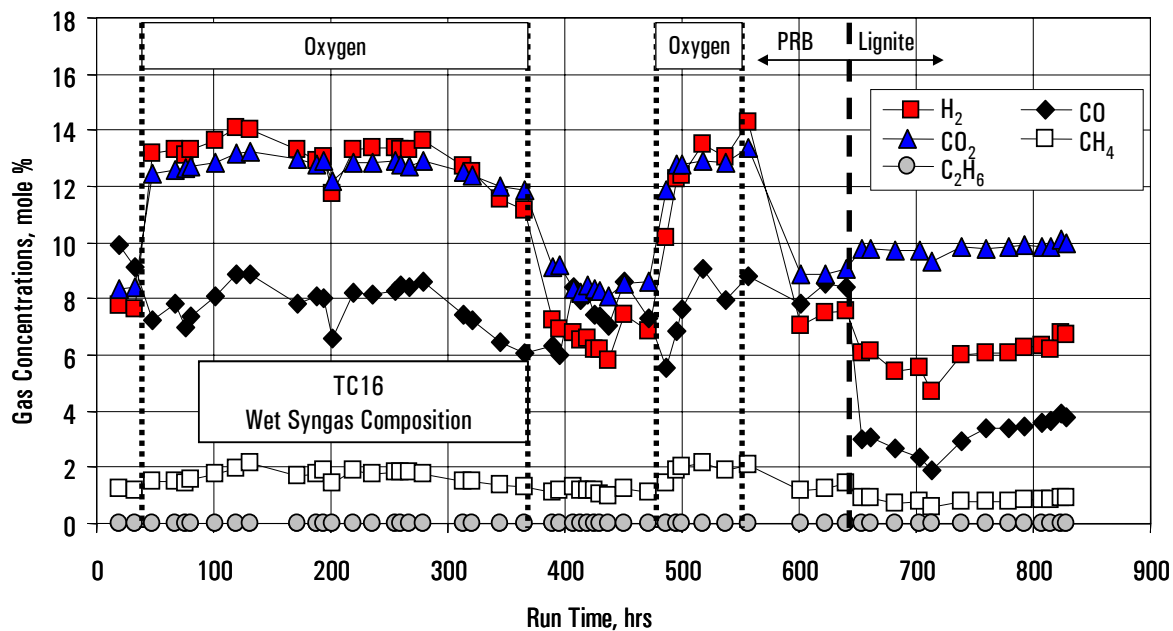


Figure 3.1-2 Wet Syngas Compositions

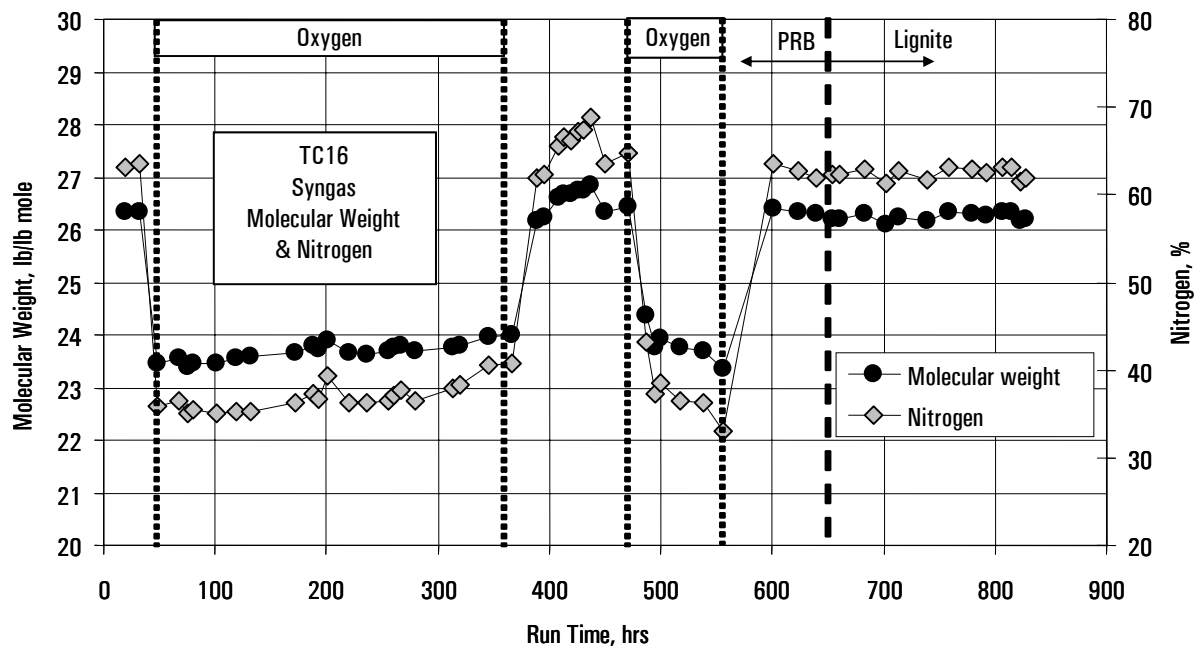


Figure 3.1-3 Wet Syngas Molecular Weight and Nitrogen Concentration

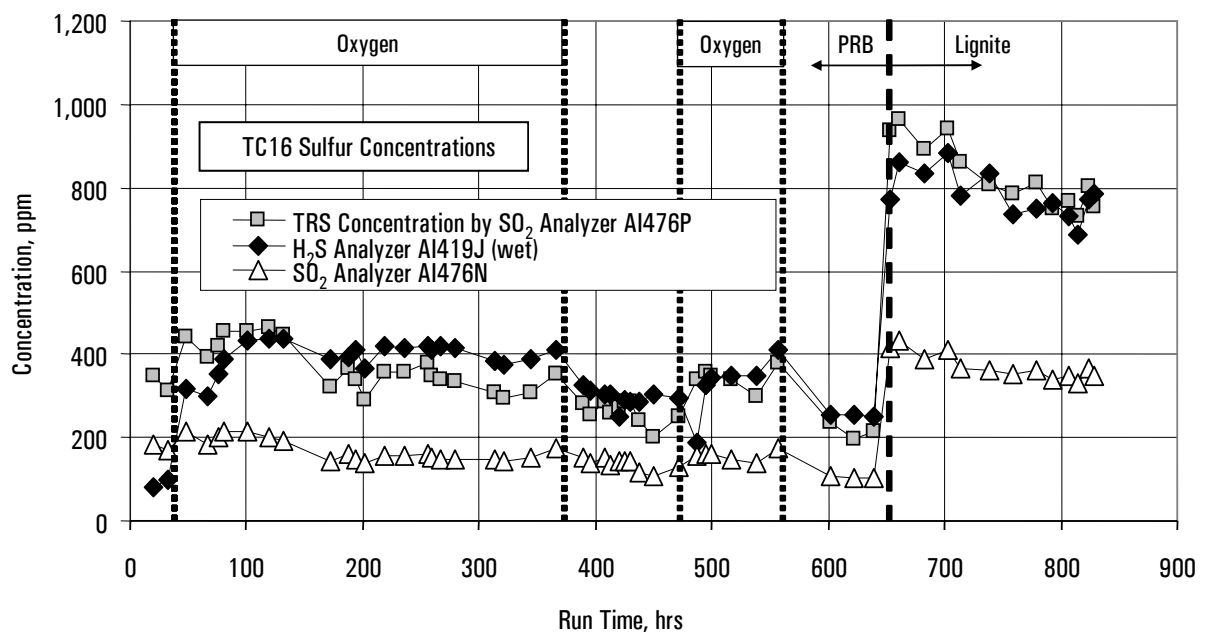


Figure 3.1-4 Sulfur Concentrations

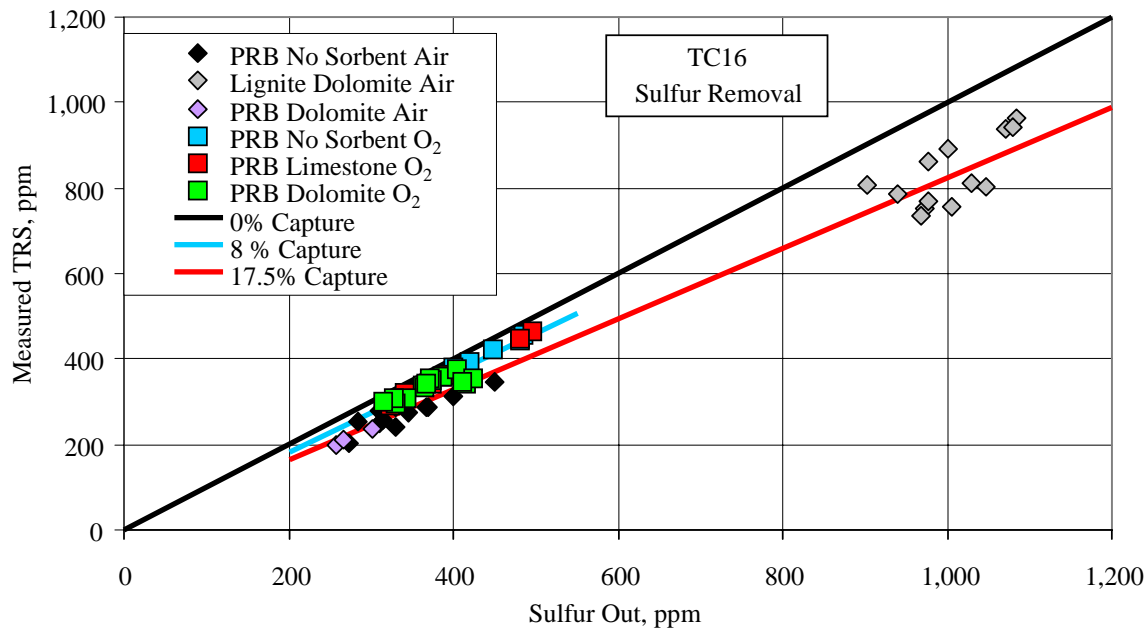


Figure 3.1-5 Sulfur Capture

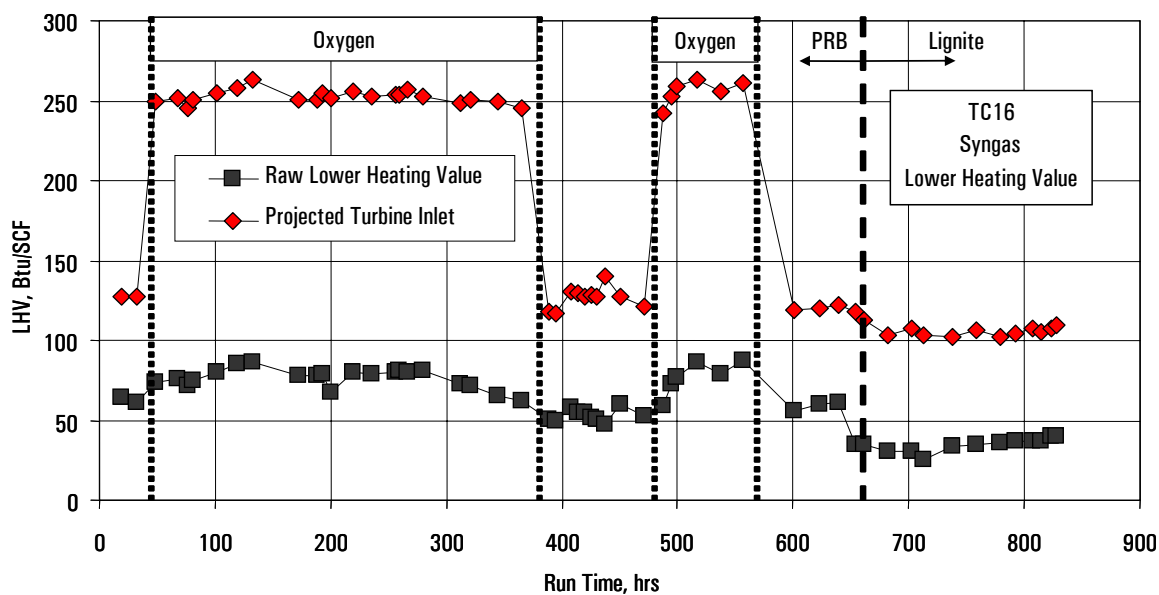


Figure 3.1-6 Syngas Lower Heating Values

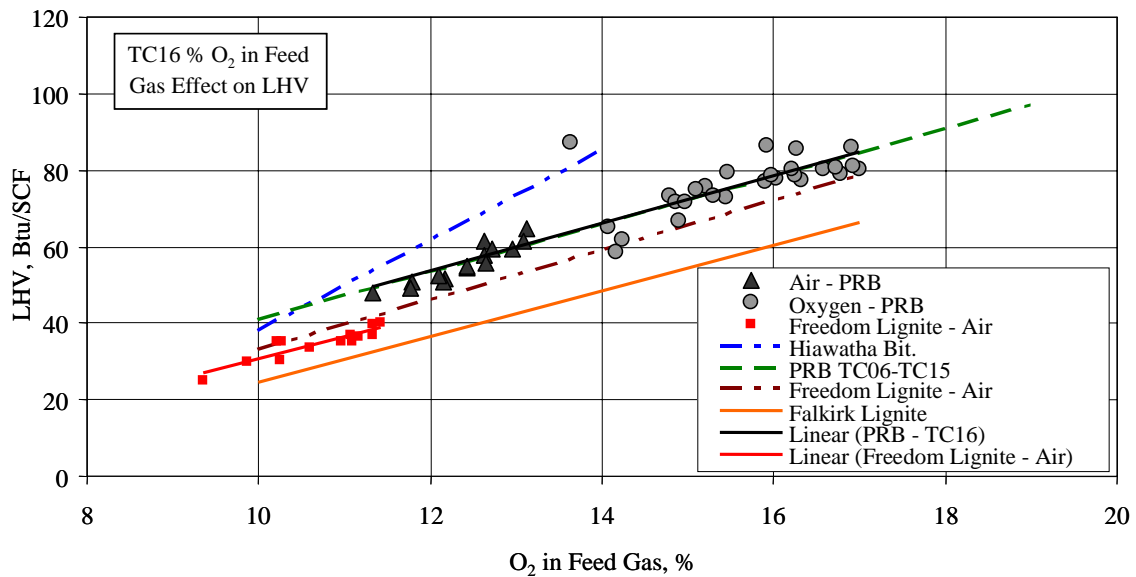


Figure 3.1-7 Raw Lower Heating Value and Overall Percent O₂ in Feed

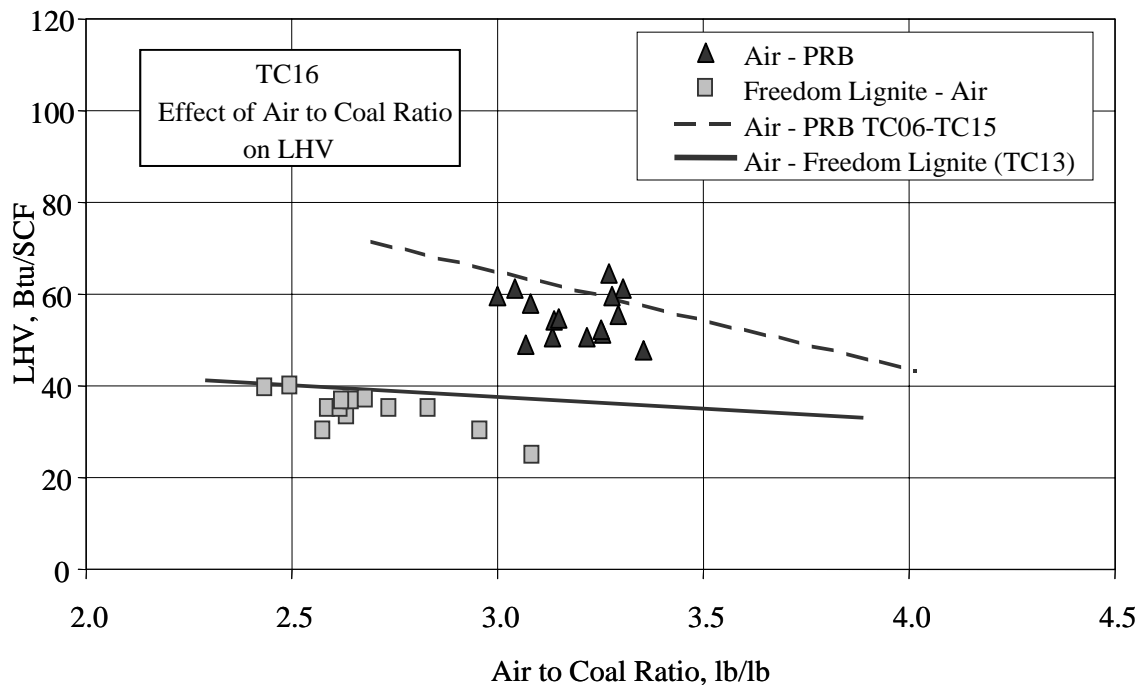


Figure 3.1-8 Raw Lower Heating Value and Air-to-Coal Ratio

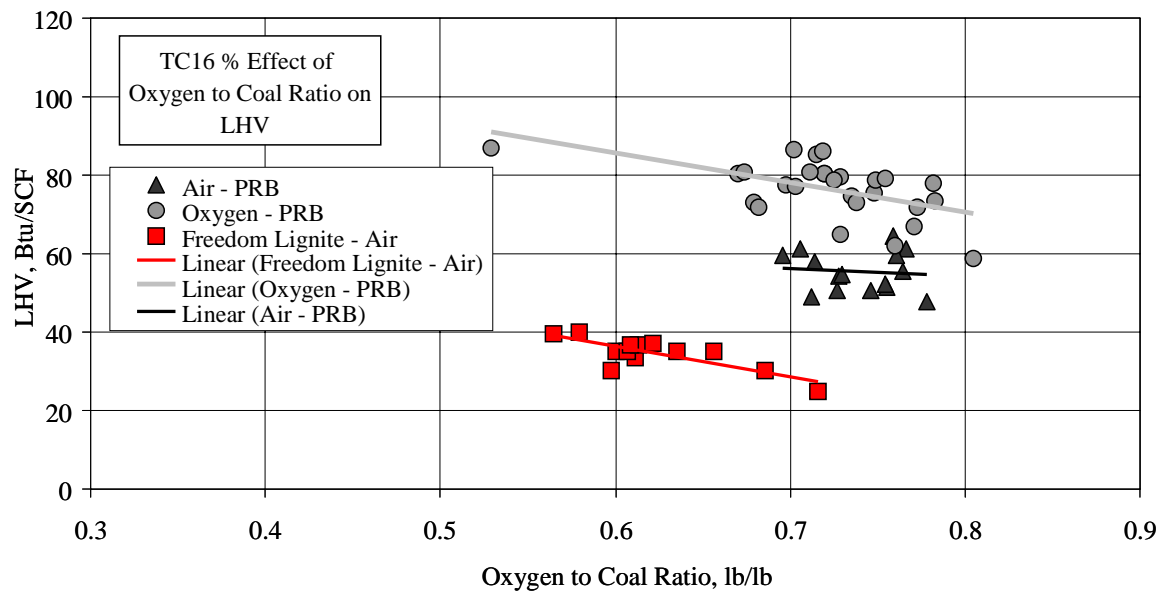


Figure 3.1-9 Raw Lower Heating Value and Oxygen to Coal Ratio

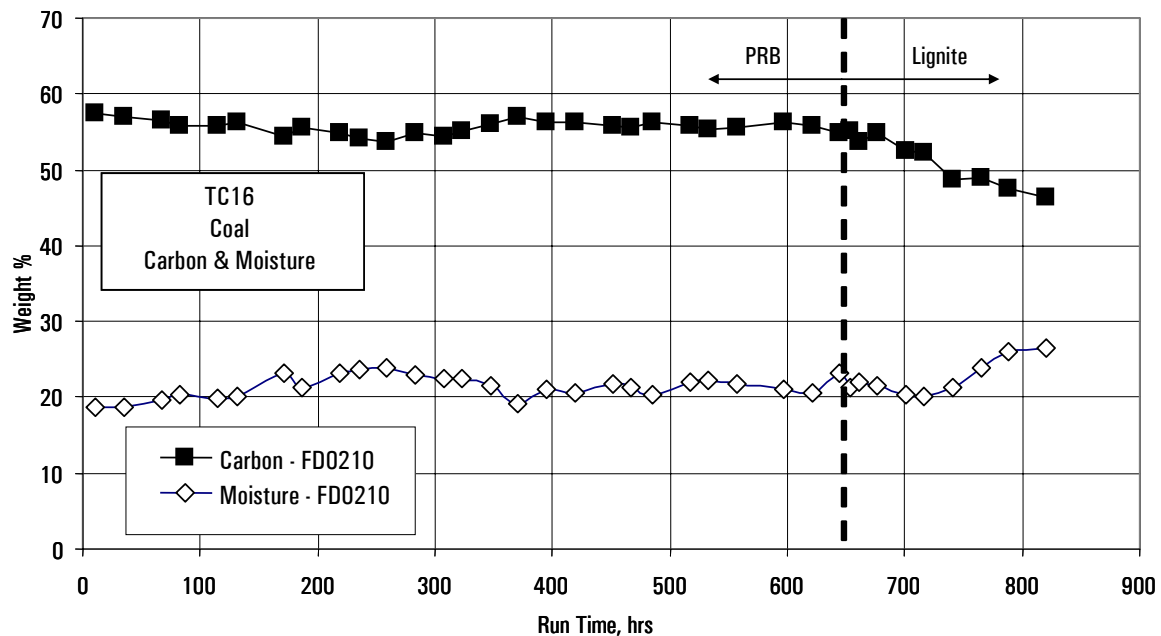


Figure 3.1-10 Coal Carbon and Moisture

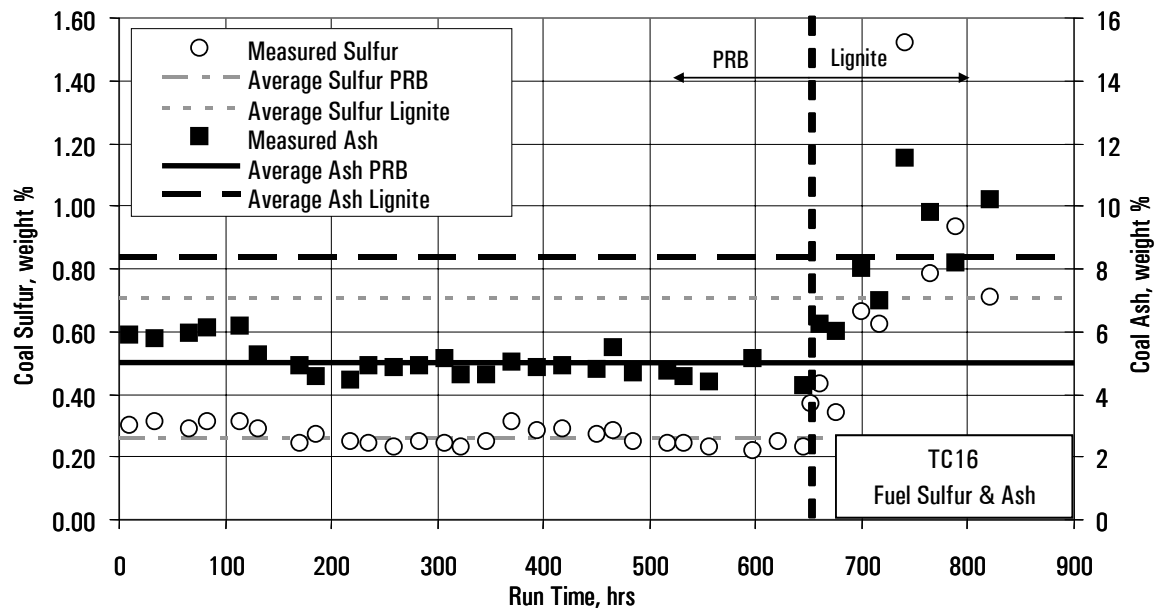


Figure 3.1-11 Coal Sulfur and Ash

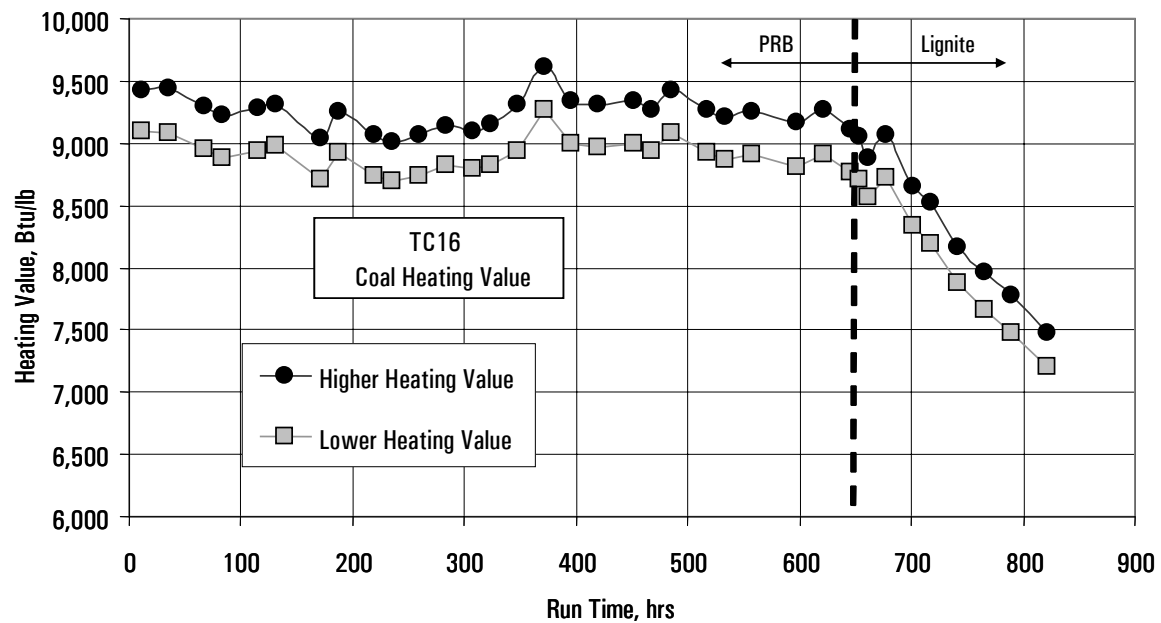


Figure 3.1-12 Coal Heating Value

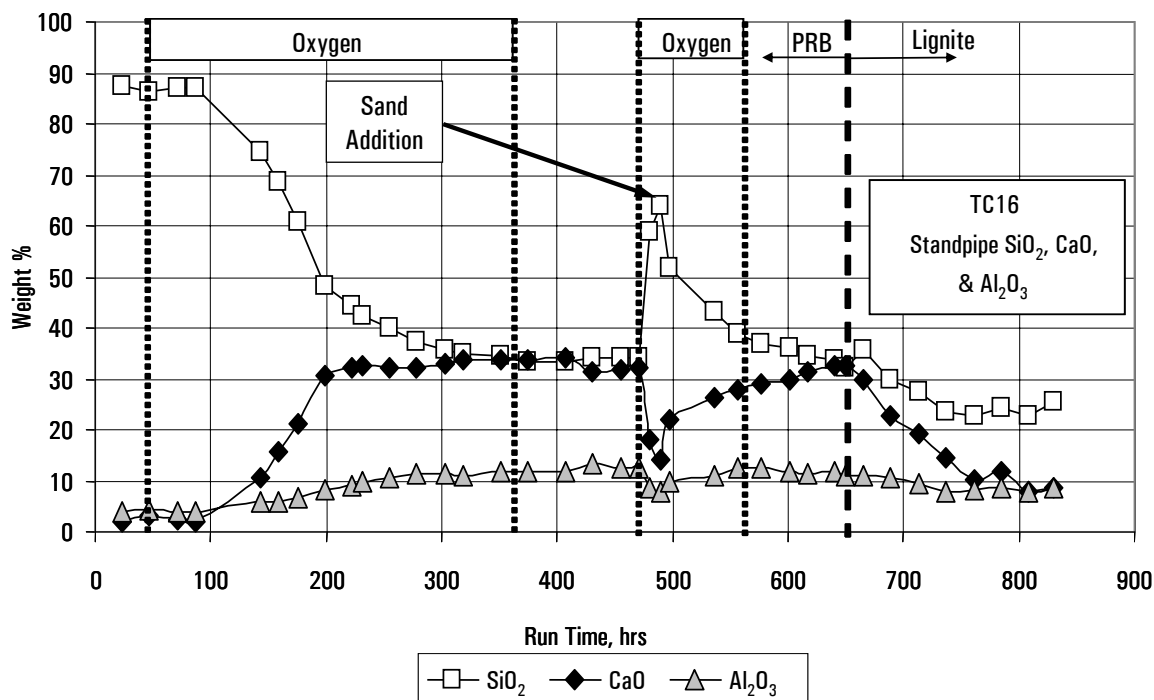


Figure 3.1-13 Standpipe Solids SiO_2 , CaO , & Al_2O_3

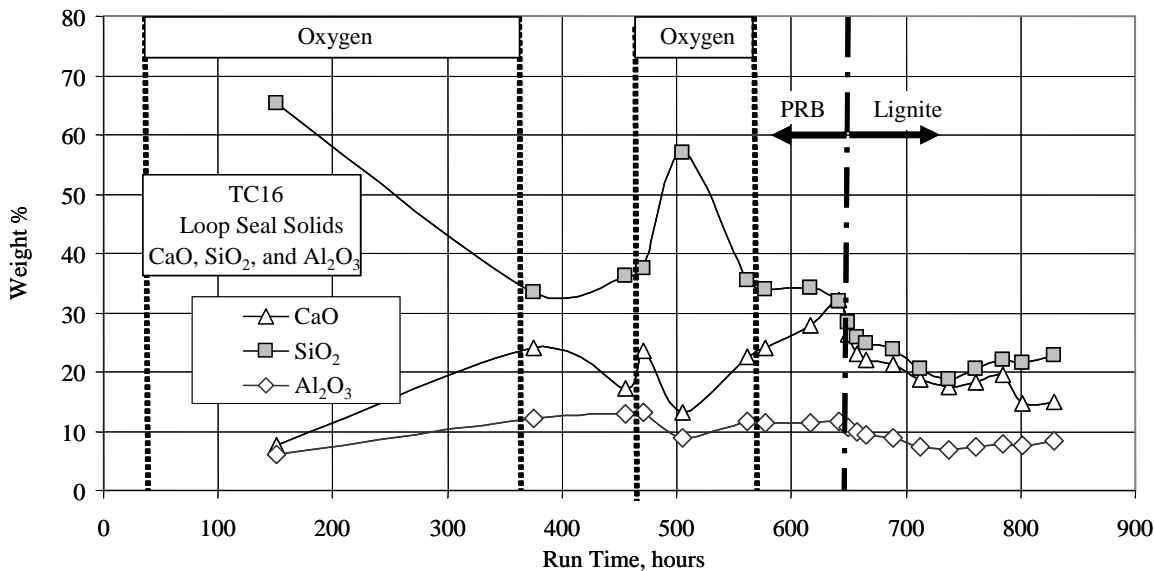


Figure 3.1-14 Loop Seal Solids SiO_2 , CaO , and Al_2O_3

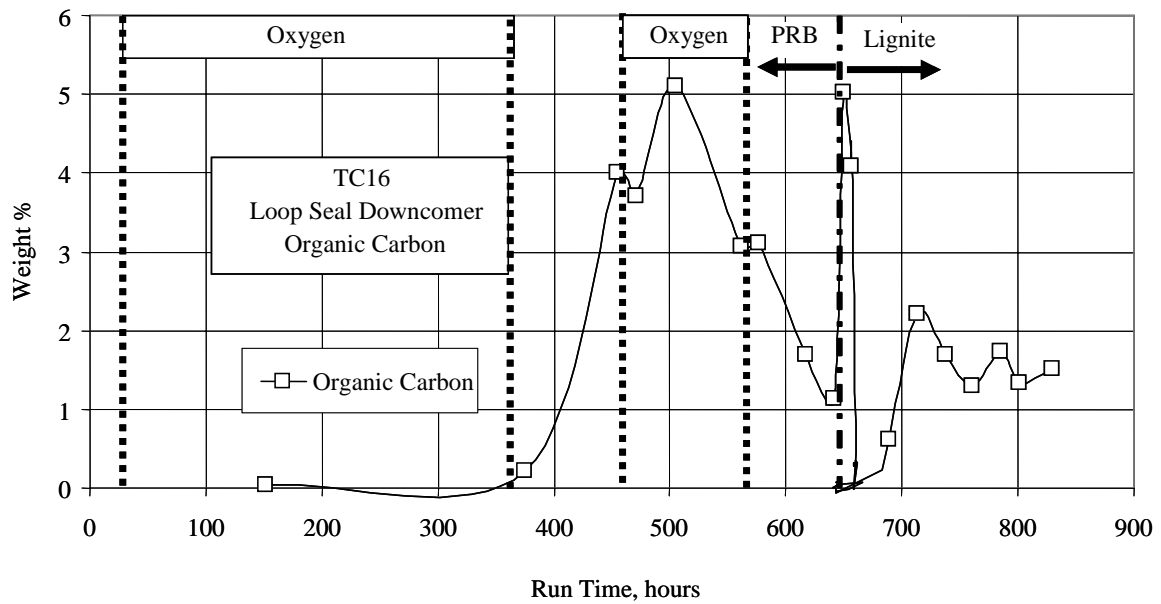


Figure 3.1-15 Loop Seal Solids Organic Carbon

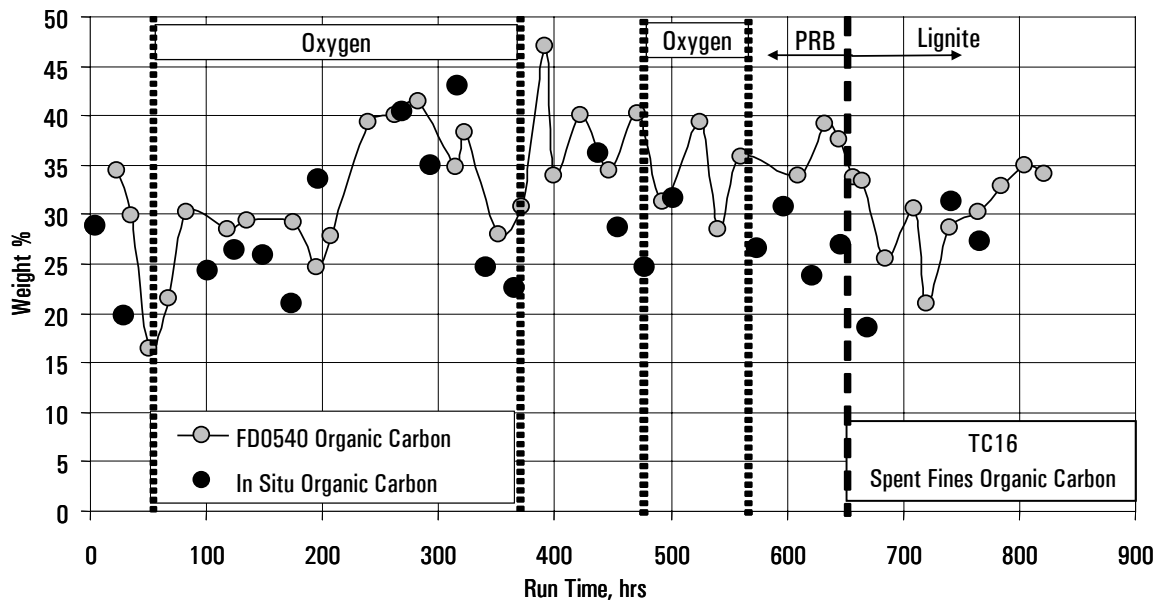


Figure 3.1-16 PCD Solids Organic Carbon

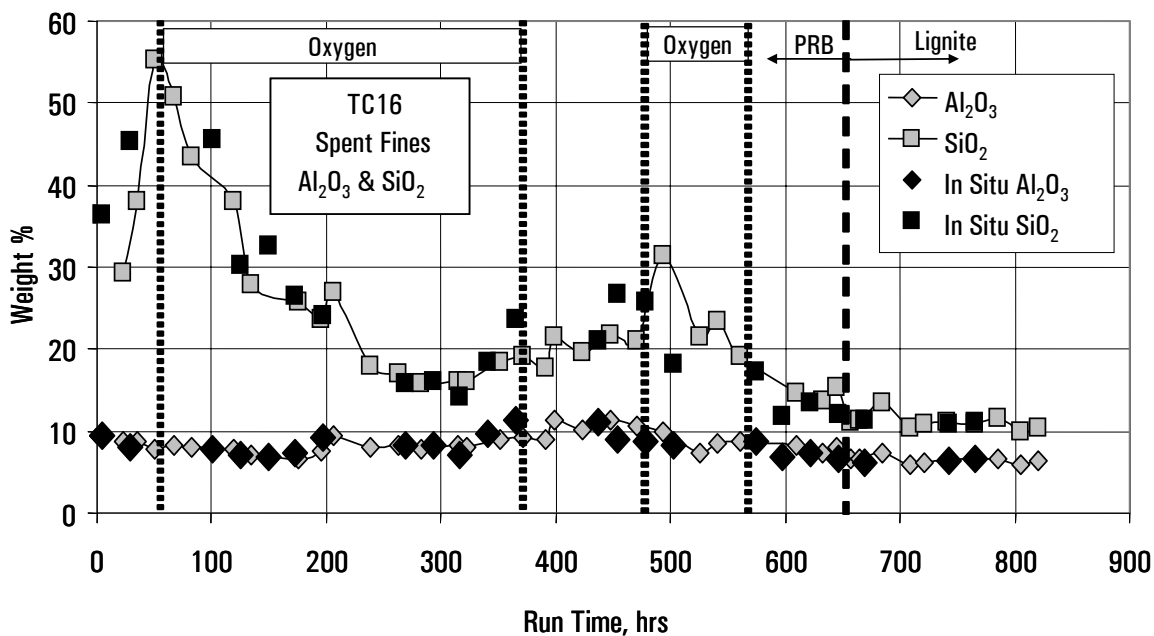


Figure 3.1-17 PCD Solids Silica and Alumina

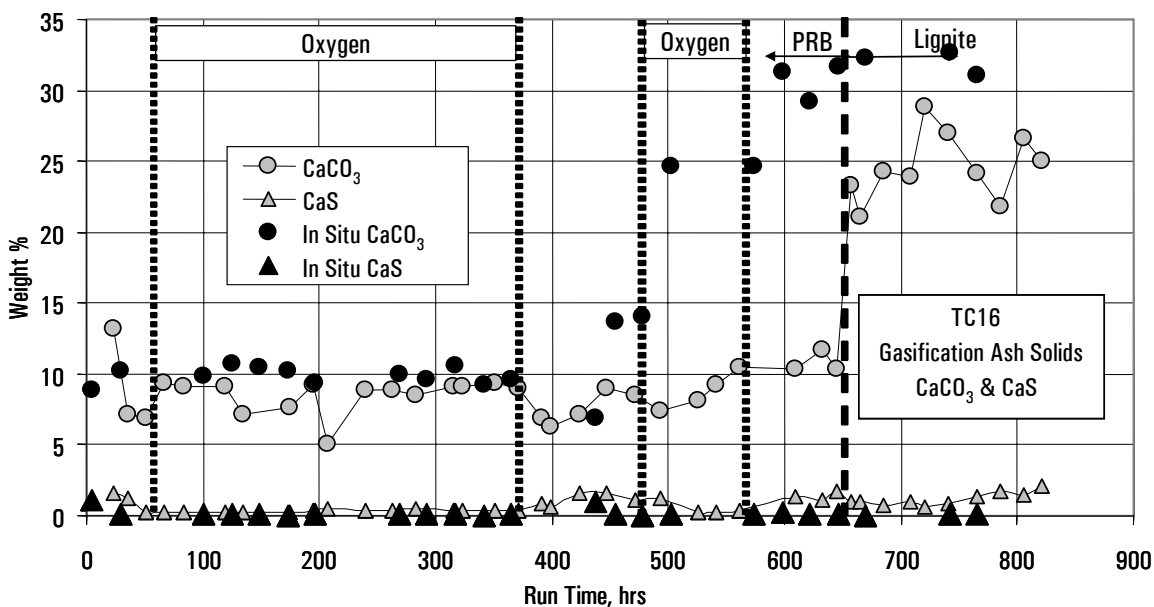


Figure 3.1-18 PCD Solids Calcium Carbonate and Calcium Sulfide

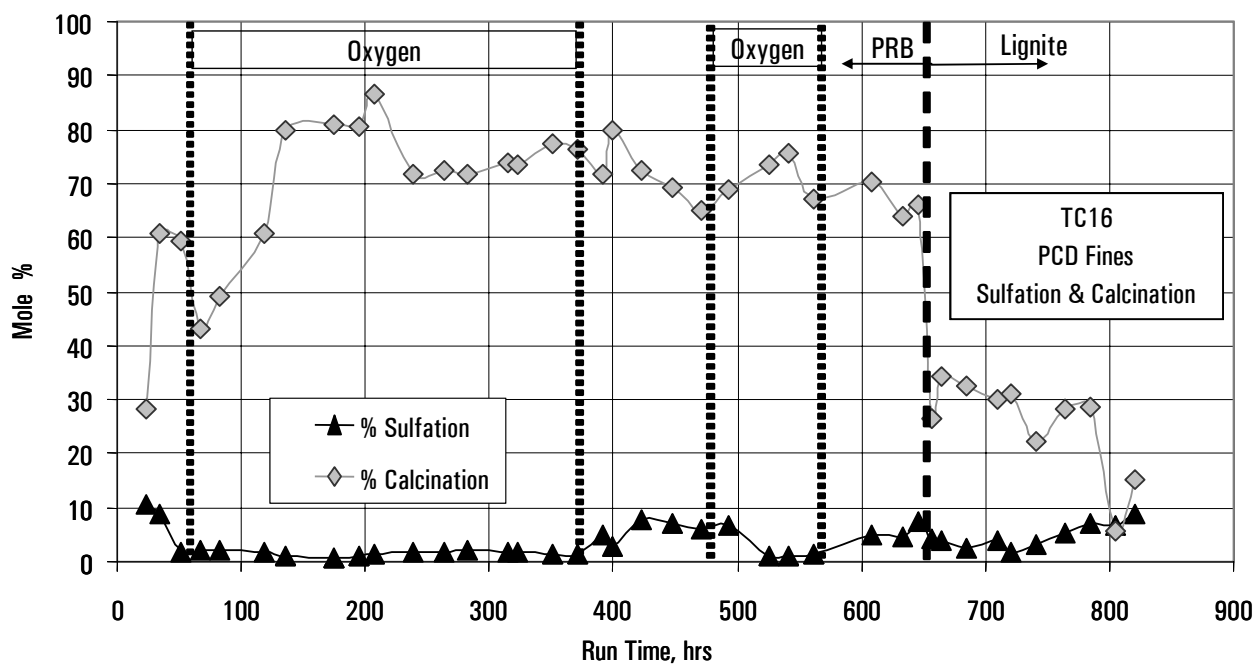


Figure 3.1-19 PCD Solids Calcination and Sulfation

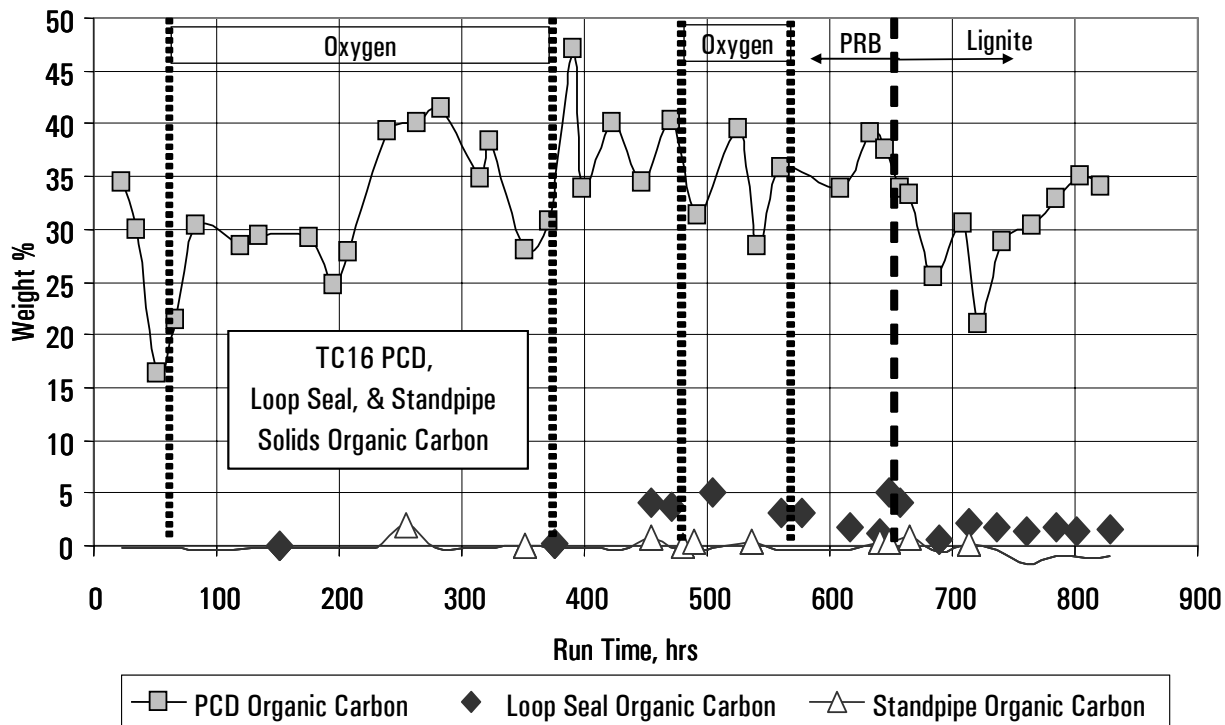


Figure 3.1-20 Standpipe, Loop Seal, and PCD Solids Organic Carbon Content

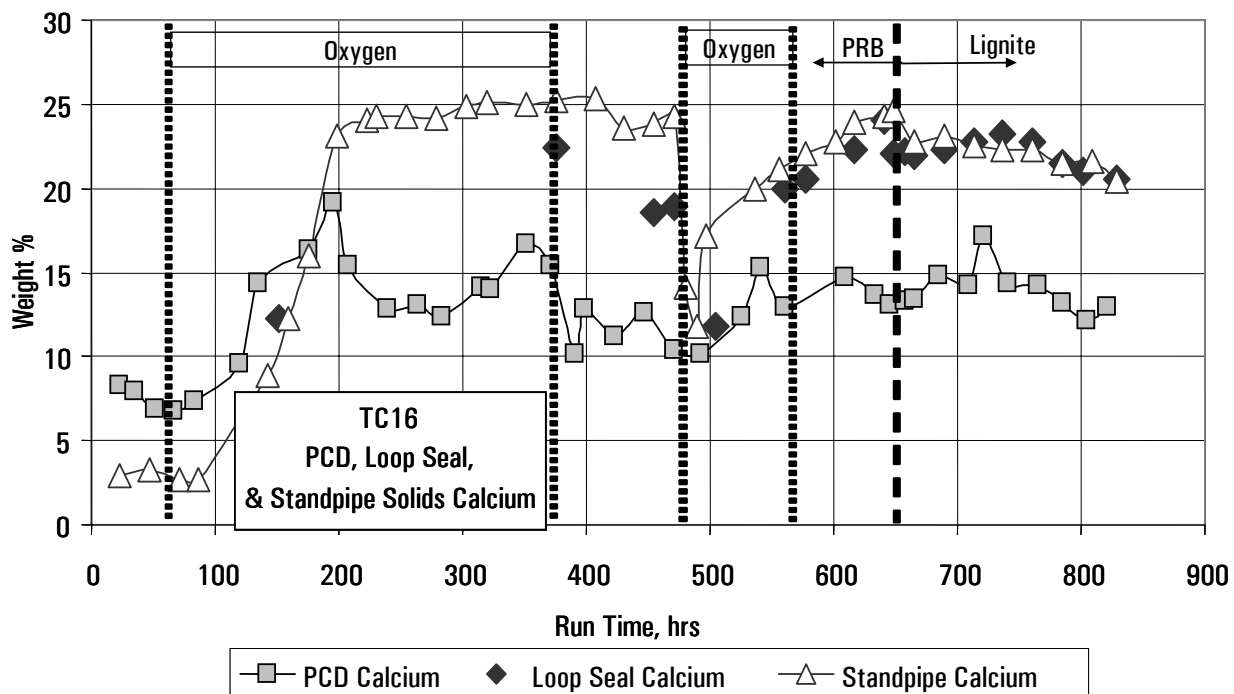


Figure 3.1-21 Gasifier and PCD Solids Calcium

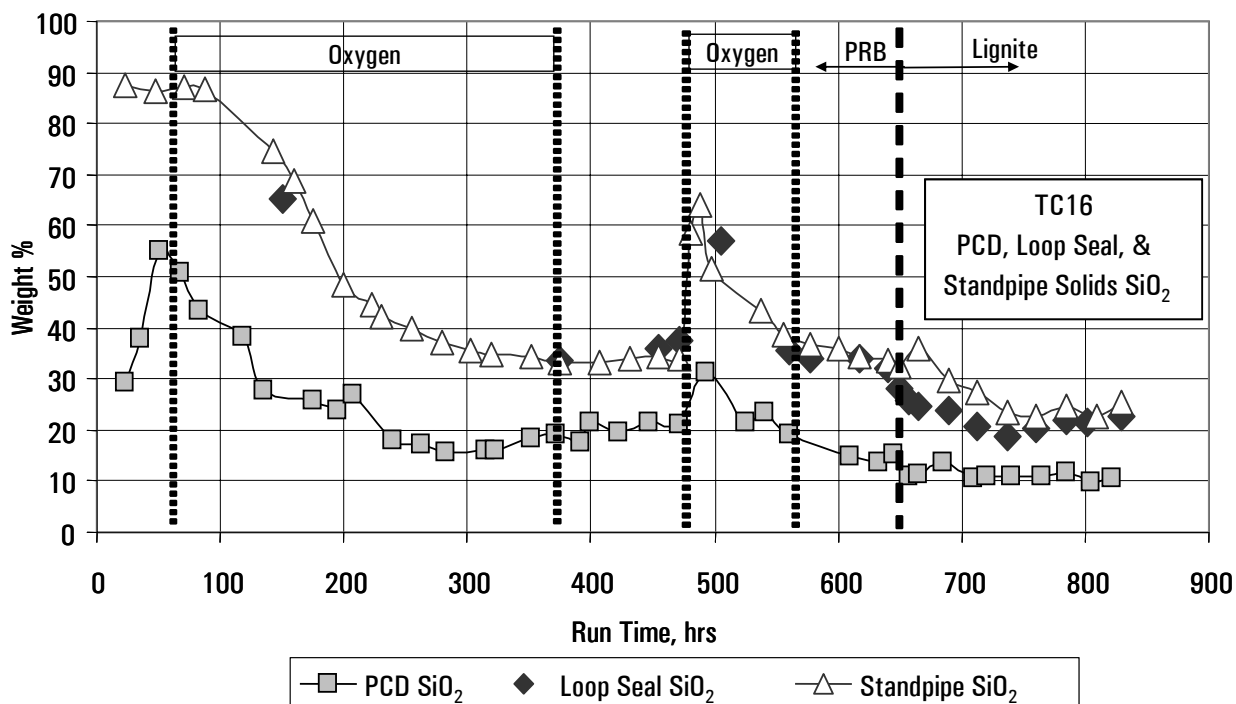


Figure 3.1-22 Gasifier and PCD Solids Silica

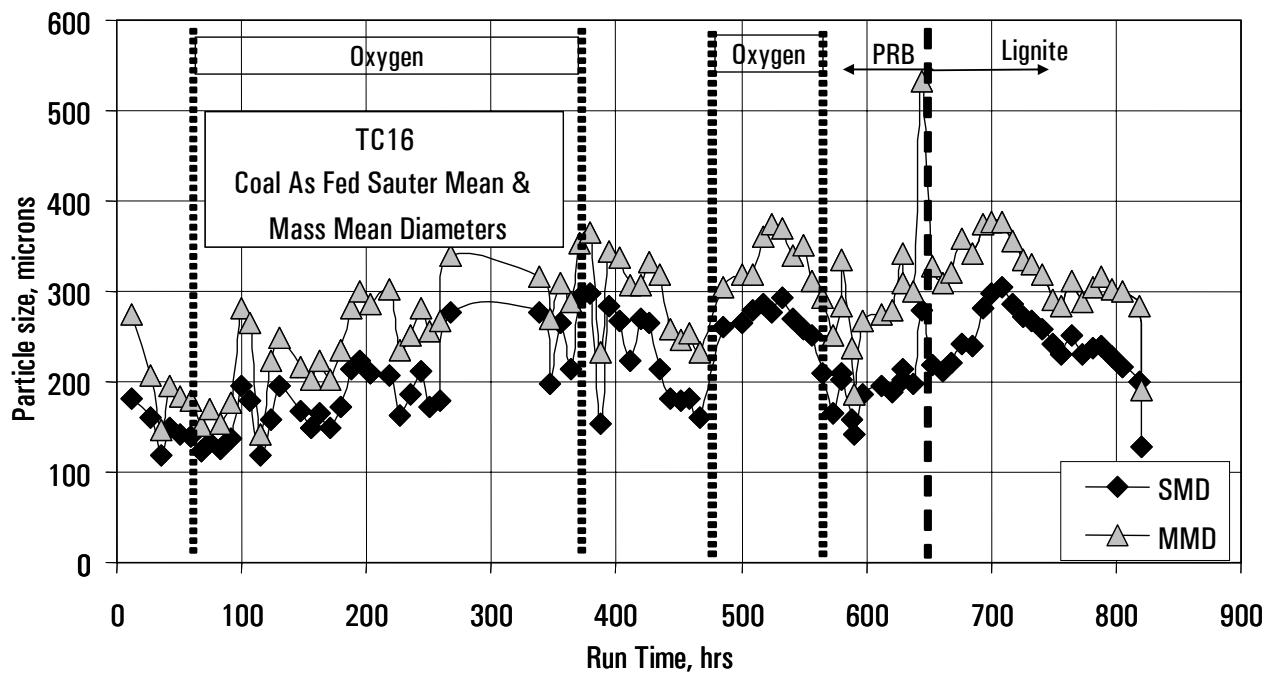


Figure 3.1-23 Coal Particle Size

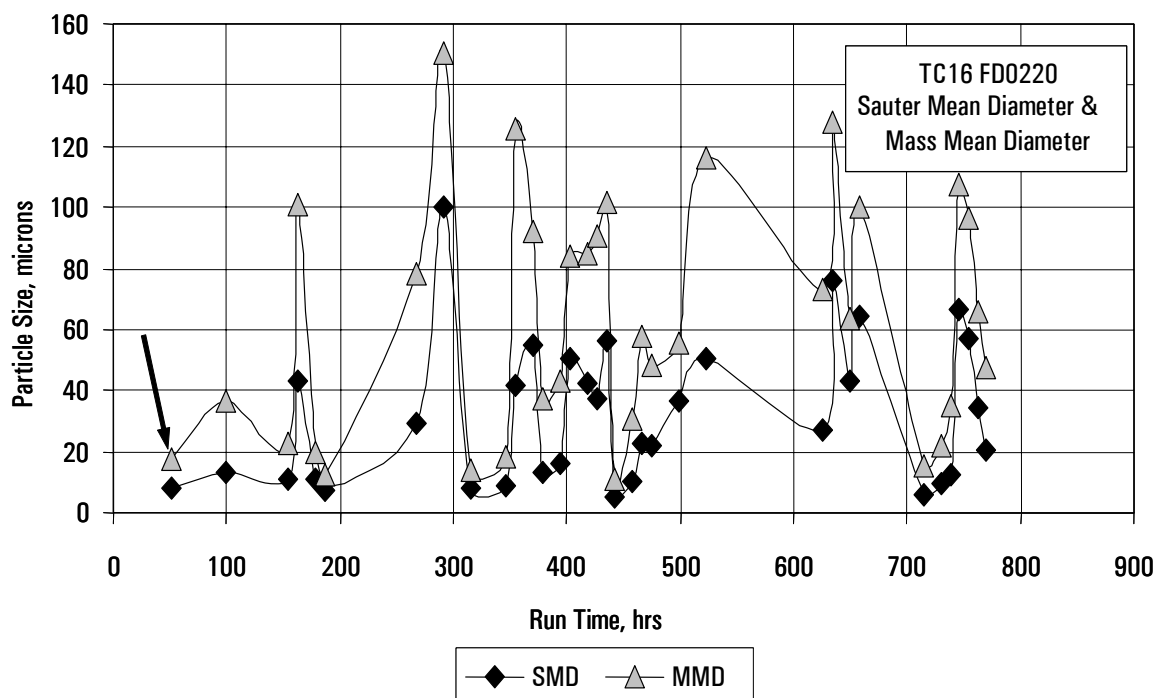


Figure 3.1-24 Sorbent Particle Size

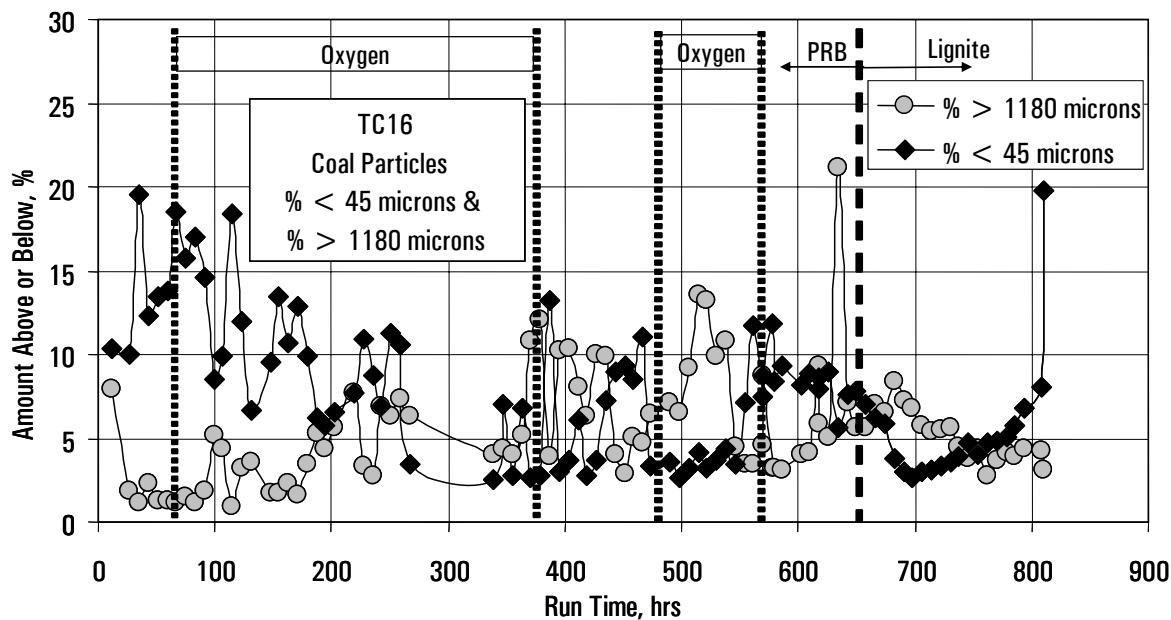


Figure 3.1-25 Percent Coal Fines and Oversize

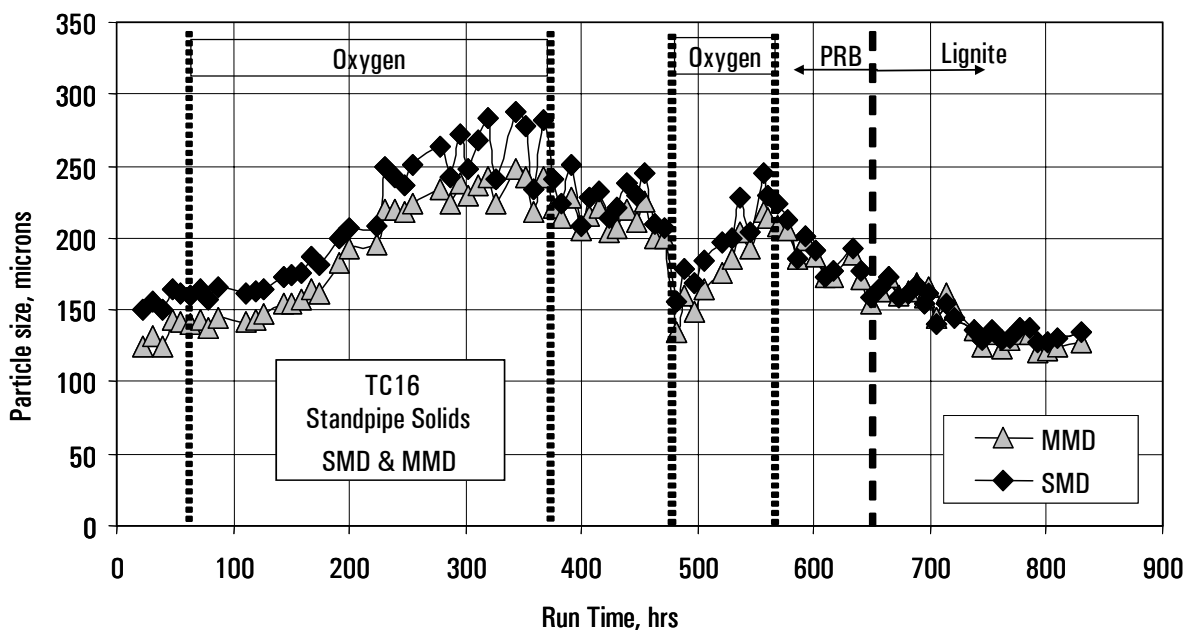


Figure 3.1-26 Standpipe Solids Particle Size

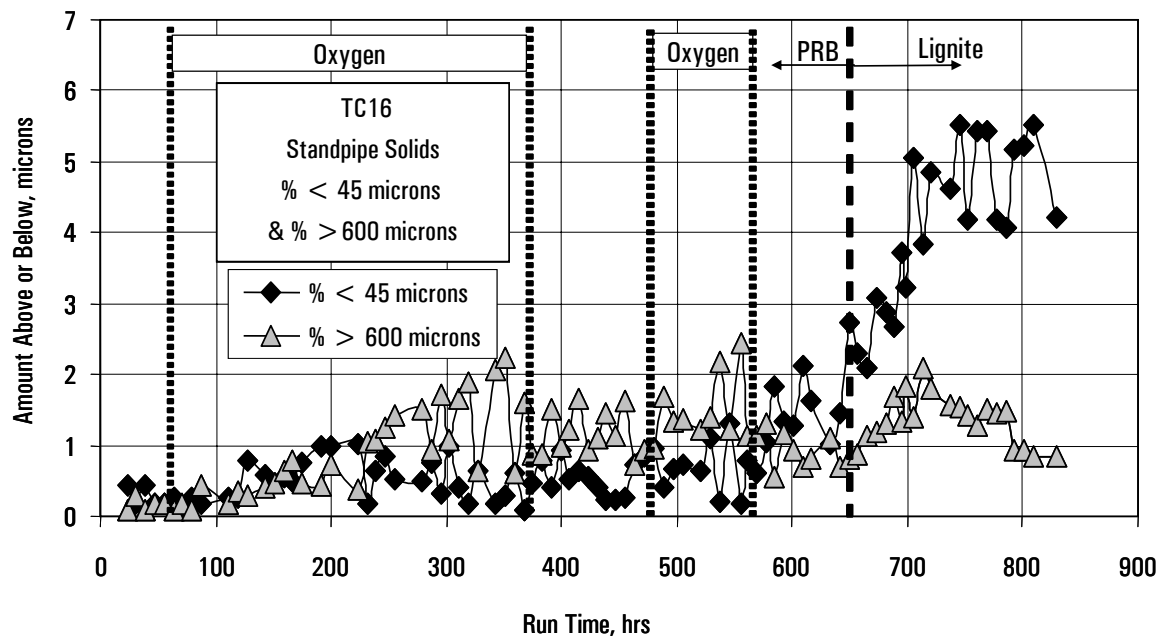


Figure 3.1-27 Standpipe Solids Fine and Coarse Particles

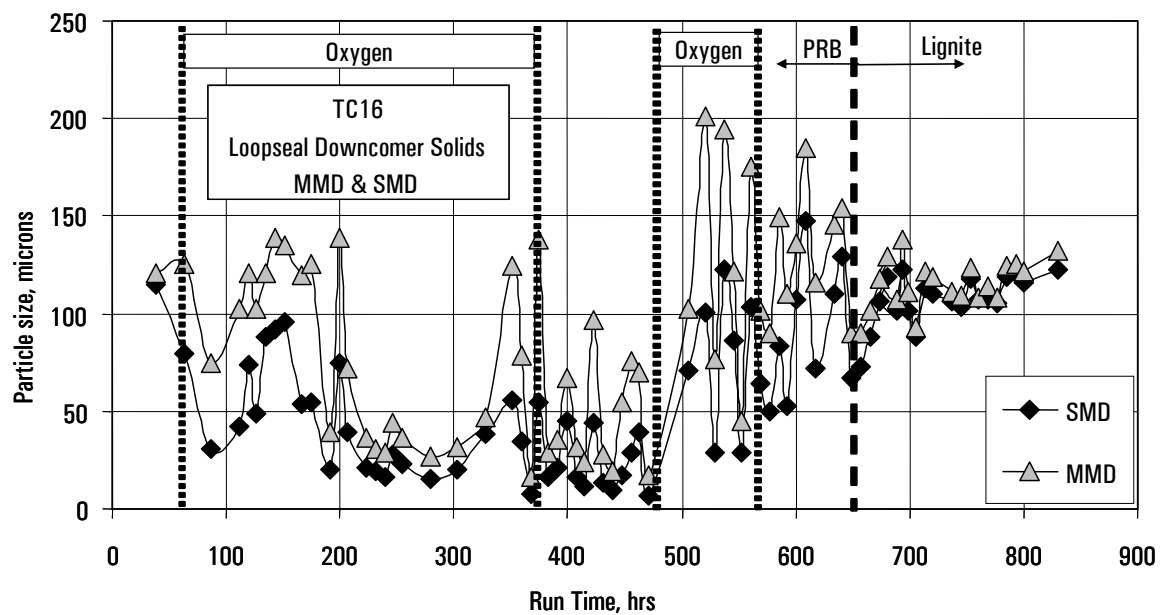


Figure 3.1-28 Loop Seal Solids Particle Sizes

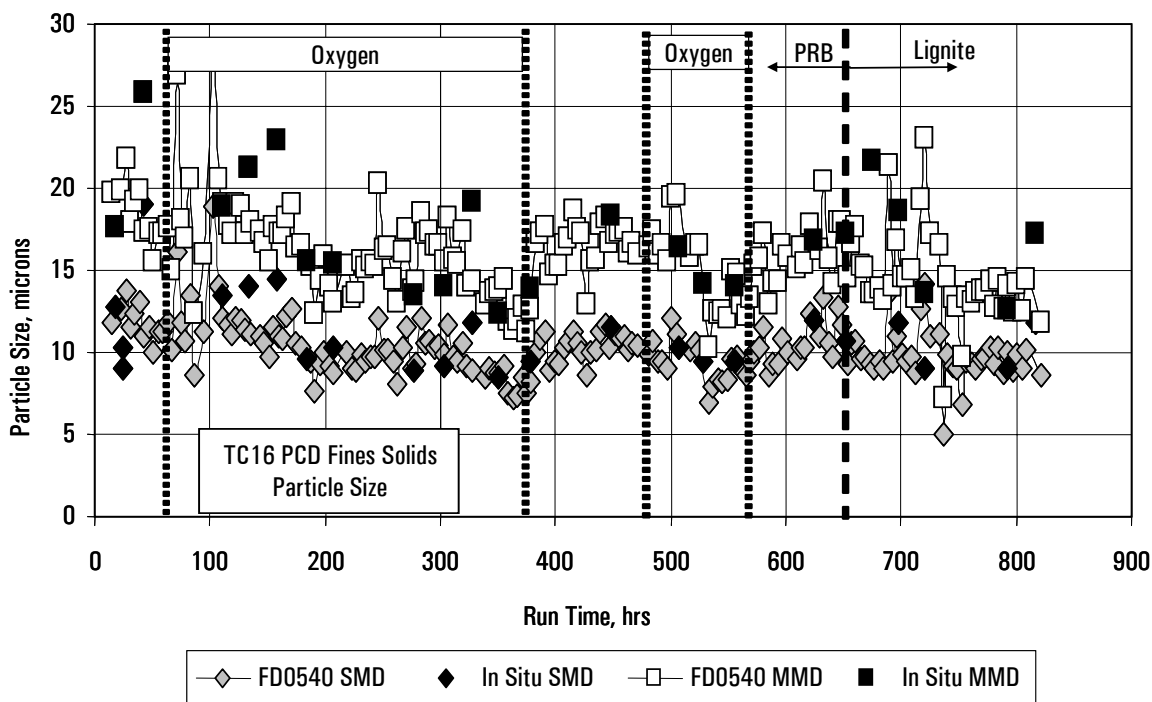


Figure 3.1-29 PCD Solids Particle Sizes

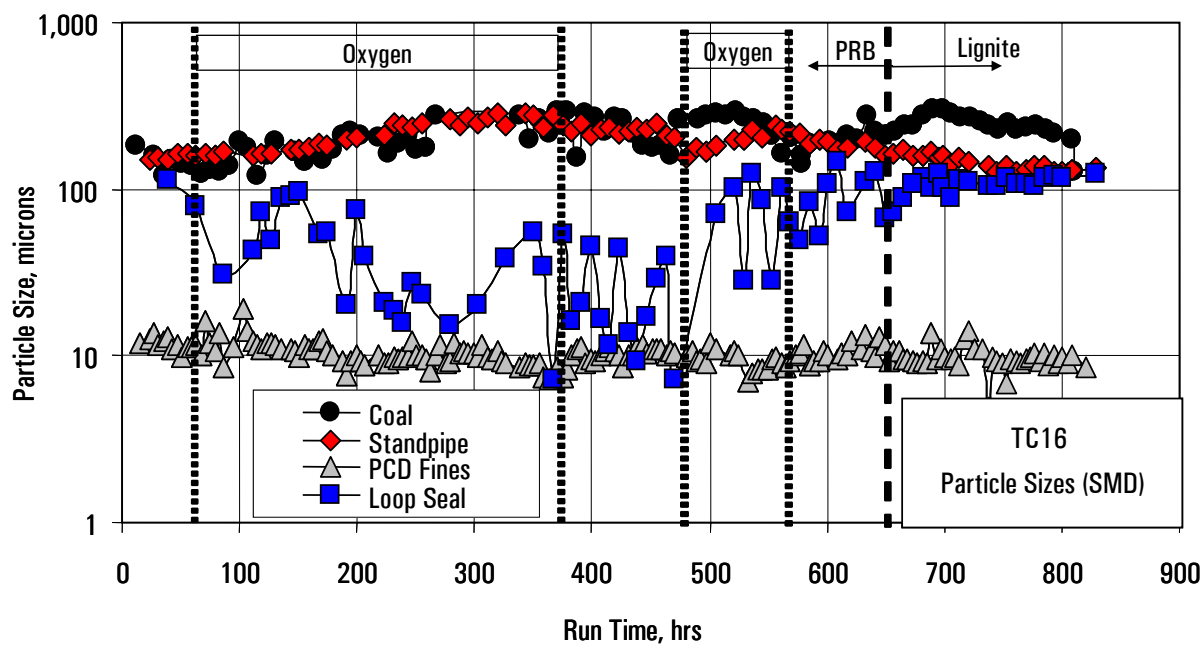


Figure 3.1-30 Particle Size Distribution

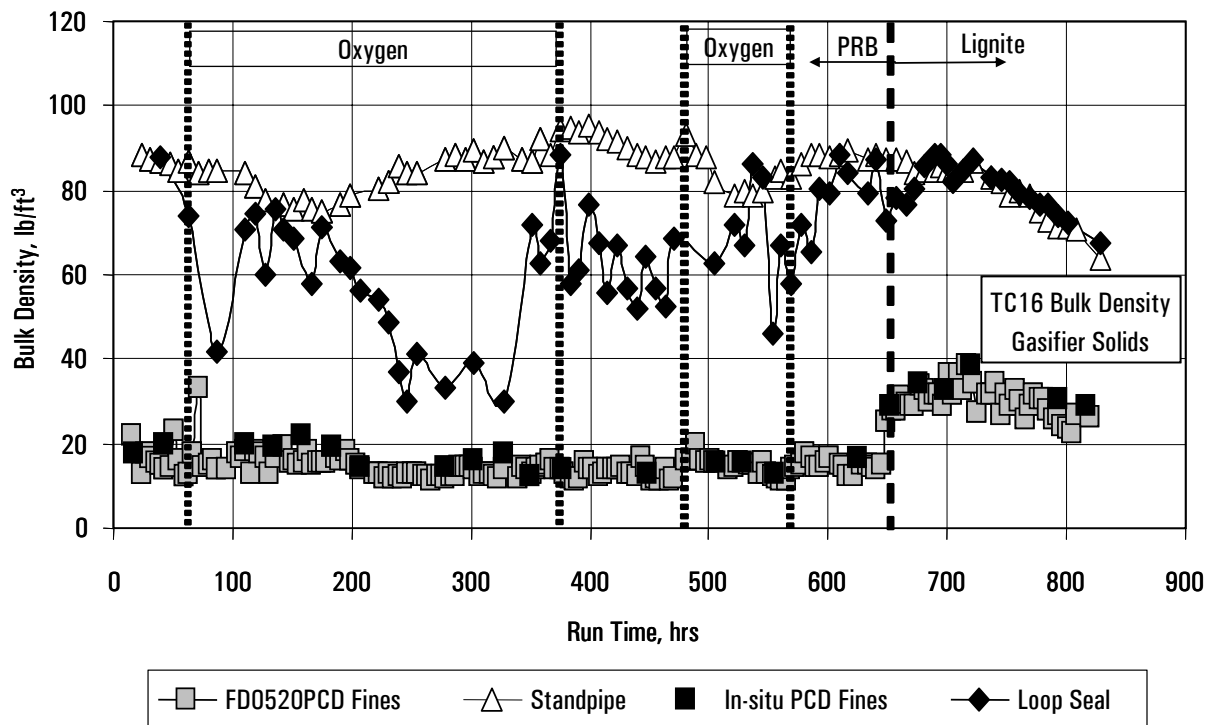


Figure 3.1-31 Gasifier and PCD Solids Bulk Density

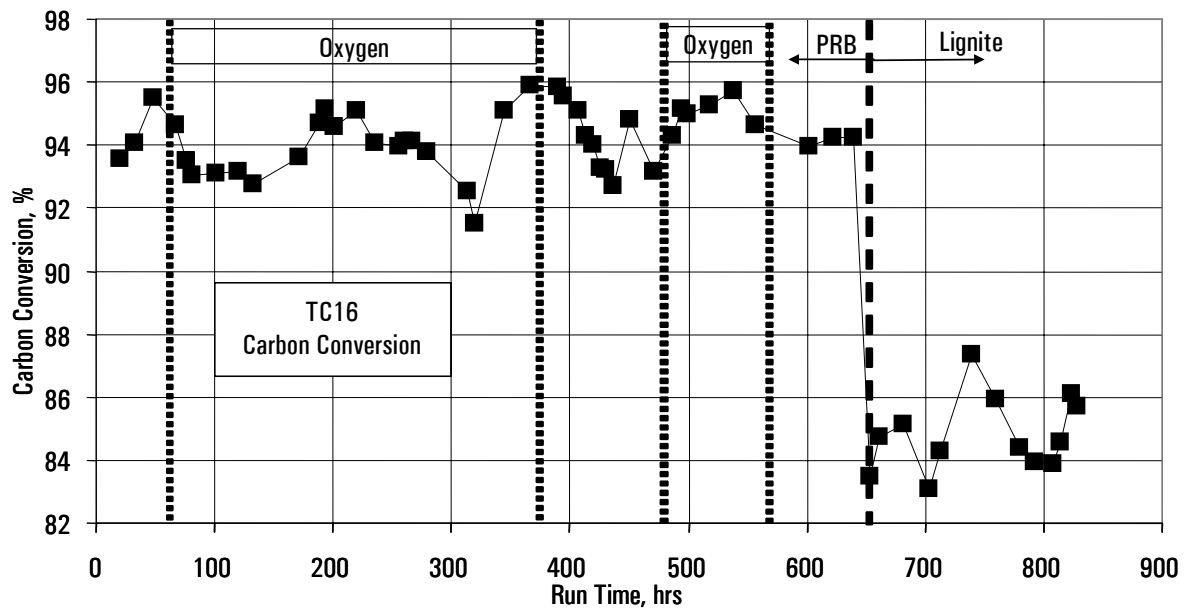


Figure 3.1-32 Carbon Conversion

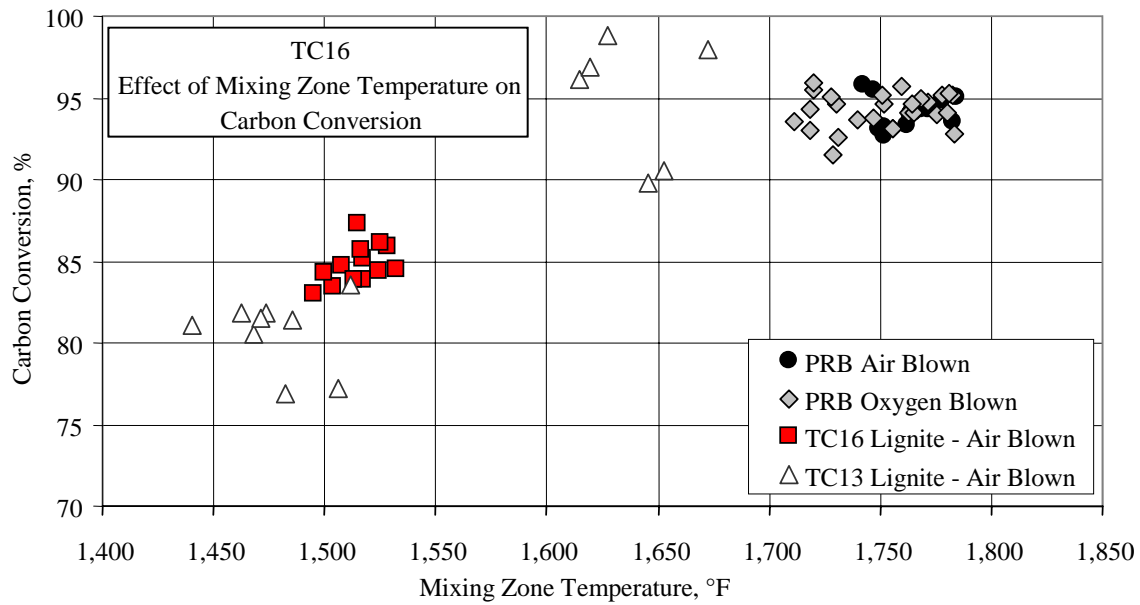


Figure 3.1-33 Carbon Conversion and Mixing Zone Temperature

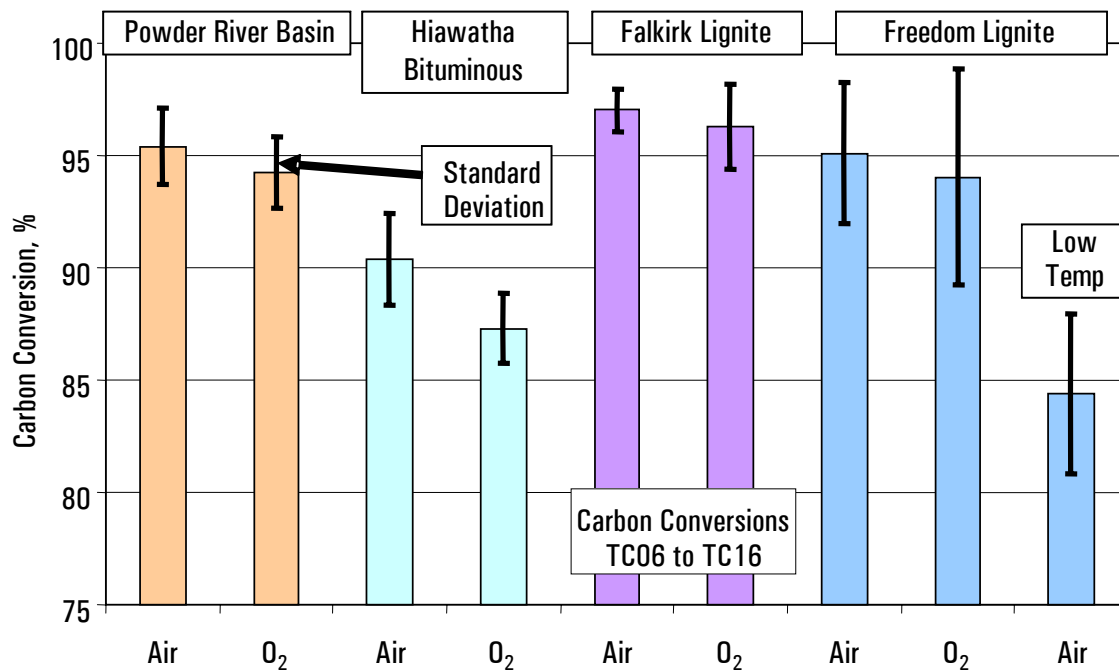


Figure 3.1-34 Carbon Conversion of Four Coals

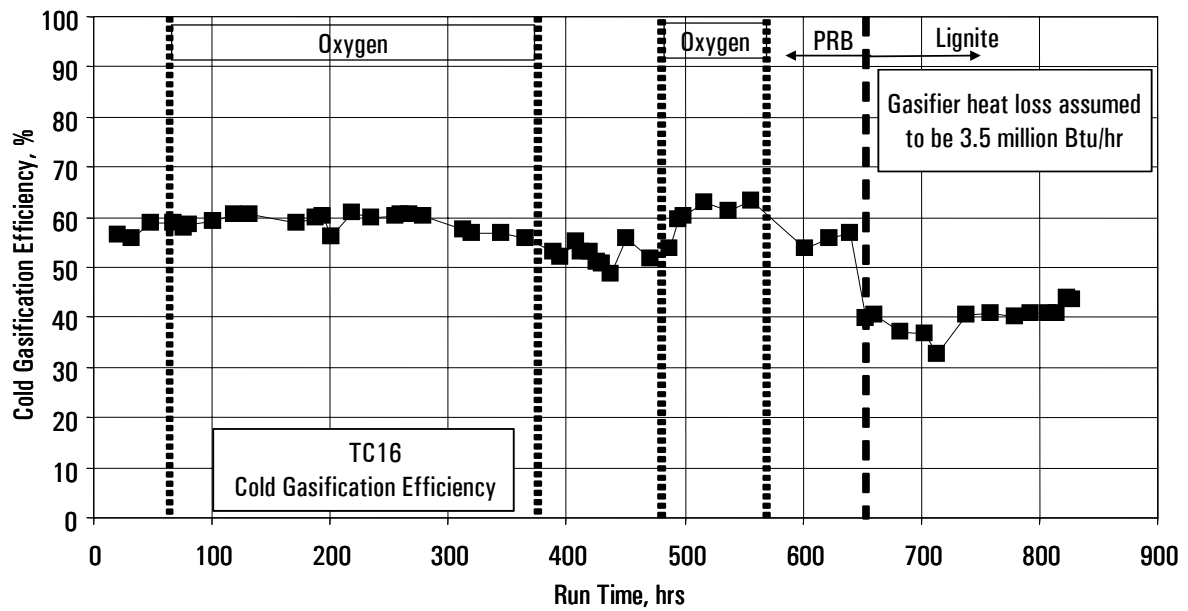


Figure 3.1-35 Cold Gasification Efficiency

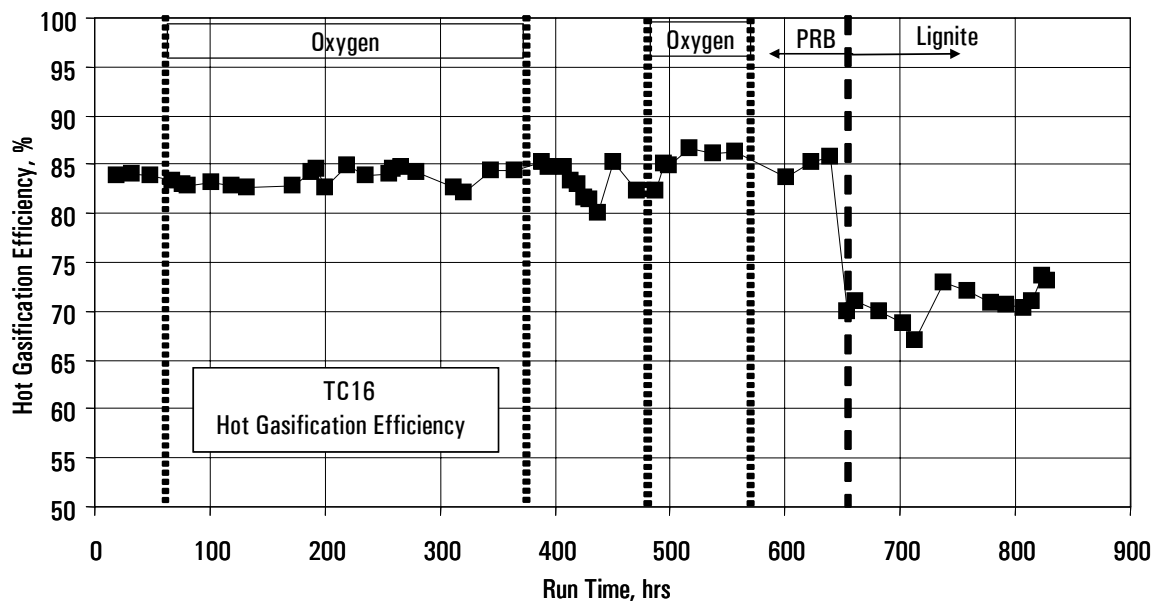


Figure 3.1-36 Hot Gasification Efficiency

3.2 PARTICULATE CONTROL DEVICE

3.2.1 Overview

Several major advances were made in operation of the particulate control device (PCD) during TC16. This was the first test run in which the failure of a filter element did not cause a system shutdown. The failsafe above the broken filter element worked so well that the outlet loading remained below the detection limit throughout the test run. Bridging occurred during the first portion of the test run after reducing backpulse parameters in an attempt to further characterize filter element cleaning requirements. The bridged material was partially removed by controlled on line combustion, which was the first time this procedure was attempted. For most of the run, the baseline pressure drop was about 60 to 120 inH₂O with higher pressure drop seen during bridging. The inlet temperature was about 650°F to 800°F, and the face velocity was maintained at 3 to 4 ft/min during steady state operations.

Seventy-two iron aluminide (FEAL) filter elements were tested to evaluate their long term performance. Several of these filter elements have been exposed to syngas for approximately 5,500 hours. Failsafe devices installed included 24 Pall fuses, 3 CeraMem ceramic failsafes, and 46 PSDF designed failsafes. Five Siemens metal fiber failsafes were also installed above blanks to continue testing material exposure to syngas.

Outlet loading samples were collected throughout the test run. A small amount of particle penetration up to 0.4 ppmw was indicated during the first two days. Outlet mass concentrations dropped below 0.1 ppmw, the lower limit of detection, for the remainder of the test run. The rupture disc failsafe tester was successfully tested with a Pall Fuse failsafe to simulate a catastrophic filter element failure. The performance of the failsafe was evaluated and the outlet loading approximately four hours after the testing device opened was less than 0.1 ppmw, which indicated that the failsafe did not leak.

Optimization testing was performed on the backpulse parameters by varying the backpulse frequency and pressure to determine the effect on the baseline pressure drop and the lower limits of operation for adequate filter element cleaning. The backpulse pressures were varied from 150 psi to 300 psi above system pressure. It was found that long-term operation at 150 psi and 20 minutes were too low for adequate cleaning because bridging was formed under these conditions. Three pressure taps were installed to evaluate backpulse intensity at various operating conditions. Pressure changes inside a filter element, on the clean side of the bottom plenum and in the dirty side of the PCD vessel were measured and recorded using a high speed data acquisition system.

TC16 provided an opportunity to study the dust characteristics and PCD operation while gasifying both Powder River Basin (PRB) coal and high sodium North Dakota lignite. With the PRB coal, PCD operations were evaluated in both air-blown and oxygen-blown operating modes, with and without the addition of either limestone or dolomite for sulfur capture. The high sodium lignite was tested exclusively with dolomite addition, since lab tests suggested that the dolomite might help control agglomeration and deposition problems that had occurred previously with that fuel. Section 3.2 discusses the effects of the different coal types, operating modes, and sorbents on the characteristics of the solids entering the PCD. The effects of the various changes on PCD performance are also analyzed.

3.2.2 Particle Mass Concentrations

As in previous test programs, in situ particulate sampling was performed at the inlet and outlet of the PCD to quantify the particulate collection efficiency and to relate PCD performance to the dust characteristics. The samples were collected using the SRI in situ batch sampling systems described in previous reports. With PRB coal, most of the testing was done in oxygen-blown operating mode, with no sorbent, with limestone addition, and with dolomite addition. With the high sodium lignite, all of the testing was done in air-blown mode with dolomite addition.

PCD Inlet Mass Loadings. Particle mass concentrations and mass rates measured at the PCD inlet are given in the left half of Table 3.2-1, and the mass rates are plotted as a function of coal feed rate in Figure 3.2-1. As shown in the plot, the particle mass rates show a positive correlation with the coal feed rate, when each test condition is considered separately. At a given coal feed rate, the solids carryover to the PCD was higher with the high sodium lignite than it was with the PRB coal. This result is to be expected based on the higher ash content of the lignite.

PCD Outlet Mass Loadings. Particle concentrations measured at the PCD outlet are included in Table 3.2-1 and compared to other test programs in Figure 3.2-2. As in several recent test programs, a slightly elevated particulate loading was measured on the first two days after startup with the concentration falling below the lower measurement limit (0.1 ppmw) on the third day. To investigate one possible cause of the initially elevated particle loadings, leakage through the filter element seals was evaluated in a lab test fixture. The lab tests showed that there was measurable gas leakage through the fiber gaskets that were being used in the PCD, even with the test fixture tightened to the nominal torque specification of 100 in-lbs. The leakage rate increased dramatically as the applied torque was reduced. This result was significant, because a check of the remaining torque values after TC16 revealed that almost all of the bolts had torque values of 30 in-lbs or less. Based on the lab leak tests, a drop in torque of this magnitude would more than double the rate of gas leakage through the fiber gaskets.

To investigate the effect of the gasket type, the lab leakage test was repeated using a graphite sheet gasket. With the graphite gasket, the leakage was completely eliminated, even with an applied torque of only 30 in-lbs. This result suggests that the compression of the fiber gasket is not sufficient to eliminate all of the porosity between fibers, resulting in the gas leakage that was observed through the fiber gaskets. Since there is leakage of gas through the gasket, it seems to be a reasonable assumption that small particles of gasification ash could also penetrate into the gasket, and possibly pass through the gasket, at least initially. With this type of leakage, a decline in the particle loading over the first few days would be expected as the fiber gaskets become plugged with gasification ash. Not only is this behavior consistent with the observed trends in the outlet particle loadings, but examination of used gaskets has shown that they are indeed plugged with gasification ash as shown in Figure 3.2-3.

At least part of the observed gasket problem appears to be attributable to degradation of the chopped fibers that are used as filler material in the gaskets. Muffle furnace tests have shown that the chopped fibers degrade into a powder when heated to 800°F. This degradation allows the gasket to relax, resulting in reduced gasket compression and increased potential for leakage, at least until

the gasket becomes plugged with gasification ash. To overcome the loss of compression, it may be advisable to use spring type, conical washers with the bolting. Lab testing suggests that conical washers would help offset the loss of gasket compression. In muffle furnace tests at 800°F, torque retention in gasket test fixtures was at least 90 percent with the conical washers compared with only 60 to 70 percent without the washers. Based on these findings, the use of conical washers on a limited number of filter elements will be tested in future test runs.

3.2.3 Real Time Monitoring

The PCME DustAlert-90 particulate monitor, referred to as the PCME, was operational throughout TC16. As seen in previous tests, there was considerable noise in the PCME signal along with spikes that coincided with the PCD backpulsing. As in previous runs, the response of the PCME was evaluated by injecting known concentrations of gasification ash from the fluidized bed dust injection system downstream from the PCD and upstream of the PCME. The concentration of the injected solids was measured independently using the in situ particulate sampling system at the PCD outlet, which is downstream from the PCME. Figure 3.2-4 shows the PCME output during the TC16 injection tests and during previous particulate monitor evaluations. As shown in the plot, the PCME response was much greater in TC16 than it was in the earlier runs.

To investigate the cause of the increased PCME response, sampling filters collected during evaluations were examined under the microscope. This examination showed that there were many large ($> 50 \mu\text{m}$) particles being generated by the fluidized bed injection system during the test run. When the sampling filters from previous PCME evaluations that showed lower responses were examined, no particles larger than $50 \mu\text{m}$ were shown. The difference is illustrated in Figure 3.2-5, which compares the large particles generated in the TC16 tests to the particles on a sampling filter from TC14 in which the fluidized bed apparently behaved much differently. Without the large particles present, a response of only 1.8 percent was obtained with an injected concentration of 2 ppmw in TC14. With the large particles present, a response of 8.3 percent was obtained with an injected concentration of only 1 ppmw in TC16. This comparison clearly shows that the sensitivity of the PCME was strongly dependent on the presence of large particles. In situations where there are relatively small PCD leaks, these large particles may not be present, and the PCME may not be able to detect the leak.

3.2.4 PCD Solids Analysis

PCD performance can be influenced by changes in the characteristics of the solids being collected in the PCD. In particular, the properties and composition of the solids can affect the PCD pressure drop, cleaning requirements, and bridging tendency. Important characteristics of the solids include particle size distribution, bulk density, true density, porosity, surface area, composition, and flow resistance. The effect of each of these parameters is considered in analyzing the performance of the PCD.

3.2.4.1 Particle Size Distributions

A Microtrac X-100 particle size analyzer was used to measure the particle size distributions of the in situ particulate samples collected at the PCD inlet and the PCD hopper samples used for the laboratory drag measurements.

In Situ Samples. Figure 3.2-6 compares differential percent mass particle size distributions measured on the PCD inlet in situ samples for PRB coal with both air and oxygen as the oxidant. Although there were differences in the mass concentrations of particles of all sizes (Table 3.2-1) associated with the two operating modes, this display indicates that the relative numbers of particles in different size ranges were constant. Therefore, the dustcake flow resistance produced by the two modes is not expected to be different because of changes in numbers of small particles generated.

The size distributions measured at the PCD inlet with PRB coal with different sorbents added (or not added) to the gasifier are compared in Figure 3.2-7. Once again, very little difference that would affect PCD operation was observed in these data. There were some variations for the largest size particles but caution should be taken in drawing conclusions from the data at the extreme ends of the distributions because the differences are not significant.

The particle sizes related to PRB coal versus high sodium lignite are shown in Figure 3.2-8. There were no directly comparable test conditions obtained, so a range of conditions were analyzed. Considering all the factors that were varying between these three test conditions, there were no clear differences that could be attributed to the operating parameters.

Hopper Samples. Any significant particle size differences between the selected PCD hopper samples and the in situ samples could be a reason for concern, because it could indicate that the hopper samples that were used for the lab drag measurements were not representative of the gasification ash entering the PCD. If the hopper samples that were used for the drag measurements contained more fine particles than the dust typically collected in the PCD, then the lab measurements may give higher drag than that observed for the actual PCD. The differential mass percentage distributions for the in situ and hopper samples used for lab drag measurements are compared in Figure 3.2-9. The size distributions of the hopper samples show good agreement with those of the in situ samples, so the samples selected for the drag measurements were representative in terms of particle size.

3.2.4.2 Dustcake and Bridging Observations

As mentioned previously in Section 2.4.2, a bridged deposit was found after the first shutdown that covered approximately one-fourth of the bottom plenum surface area. Samples of the bridged deposit and the residual dustcake were collected. Both samples were black, suggesting that the controlled burnoff may have been incomplete. It was not possible to make dustcake thickness measurements or collect samples of the top plenum cake due to the limited access through the manway.

After the final TC16 shutdown, a complete inspection and sampling of the dustcakes were performed. No bridging was observed in either plenum, but unusually thick patches of dustcake were found. The patches were about 3/8 inch thick and extended from the top of each element, covering anywhere from 25 to 100 percent of the element length. All of the elements with 100 percent coverage were interior elements. The thick patches appeared to consist of several different layers that were roughly equal in thickness. The layers were sampled separately to determine any differences in the properties or chemistry. The color of the layers varied from a light to medium gray, but all of the layers were lighter in color than was the dustcake from the first shutdown. This observation suggested that the last burnoff was more effective than the first. Because of the uneven

nature of the dustcake and the influence of the burnoff, detailed thickness measurements were not made.

3.2.4.3 Physical Properties and Chemical Compositions

This section discusses the physical properties and chemical compositions of the in situ samples collected at the PCD inlet, the PCD hopper samples used for the laboratory drag measurements, and the dustcake samples and bridged deposit.

In Situ Samples. Tables 3.2-2 and 3.2-3 give the physical properties and chemical compositions of the in situ samples collected at the PCD inlet and the composite hopper samples used for laboratory drag measurements. In the samples taken during the PRB testing, the effect of the sorbent addition was apparent in the increased levels of calcium-containing components, and increased MgO levels in the case of the dolomite addition. The effect of the sorbent on physical properties was less clear. Compared to the gasification ash produced from PRB coal without sorbent addition, the samples obtained with limestone addition had similar bulk densities, true densities, porosities, and mean particle sizes. The only significant difference was in specific surface area. The average specific surface area of the samples produced with limestone addition was $132 \text{ m}^2/\text{g}$, compared to $164 \text{ m}^2/\text{g}$ for the samples without sorbent addition. Dolomite addition had the opposite effect on average surface area ($197 \text{ m}^2/\text{g}$ with dolomite compared to $164 \text{ m}^2/\text{g}$ without sorbent). The dolomite addition also reduced the mean particle size (average of $15.3 \text{ }\mu\text{m}$ with dolomite versus $20.7 \text{ }\mu\text{m}$ without sorbent) and increased the bulk porosity (average of 90.2 percent with dolomite versus 87.1 percent without sorbent).

When the gasifier fuel was switched from PRB coal to high sodium lignite, major changes were seen in the properties and chemistry of the gasification ash entering the PCD. Most notably, the bulk porosity was reduced substantially (average of 78.1 percent for the lignite with dolomite versus 90.2 percent for the PRB with dolomite), and the surface area dropped dramatically (average of $70 \text{ m}^2/\text{g}$ for lignite with dolomite versus $197 \text{ m}^2/\text{g}$ for PRB with dolomite).

When compared to previous tests with PRB coal and no sorbent, the TC16 in situ samples produced with the same combination were similar in terms of physical properties. However, the TC16 samples contain significantly less non-carbonate carbon (average of 24.3 percent versus 32.1 to 39.4 percent in previous runs). The reduced carbon content in TC16 suggested there was an improvement in carbon conversion in the gasifier.

Figure 3.2-10 shows the specific surface area of the TC16 in situ samples plotted as a function of the carbon content, along with samples from previous PRB runs. The TC16 samples collected with dolomite addition fit the previously established trend with limestone addition (blue line), but the data for the TC16 samples with no sorbent and with limestone were inverted. The TC16 samples with limestone added were scattered about the previously established trend line for no sorbent addition (red line), while the TC16 samples with no sorbent were scattered about the previously established trend line for limestone addition. This result was expected since the effect of limestone addition in TC16 was to reduce the surface area, while the effect in TC06 and TC07D was to increase the surface area. Despite this anomaly with the limestone addition, all of the data show a trend toward increasing surface area with carbon content. This effect of carbon on surface area may

be partly responsible for the effect of carbon on drag that has been reported previously and will be discussed further in Section 3.2.4.4.

Hopper Samples. In addition to the in situ samples, Tables 3.2-2 and 3.2-3 include data for the composite PCD hopper samples that were used for laboratory drag measurements. The hopper sample from Run 1 was with PRB coal and limestone addition, while the sample from Run 2 was with high sodium lignite and dolomite addition. Except for the mean particle sizes, the hopper samples were similar to the in situ samples collected on the same day. In both cases, the hopper samples had mean particle sizes that were significantly smaller than the mean particle sizes of the in situ samples from the same day (16.9 versus 23.0 μm for the PRB + limestone sample and 12.9 versus 17.3 μm for the lignite + dolomite sample). The reason for this difference is unknown, but it should not affect the laboratory drag measurements, since drag is measured as a function of particle size, and the hopper samples were representative in all other aspects.

Dustcake and Bridged Deposit Samples. Tables 3.2-4 and 3.2-5 give the physical properties and chemical composition of the bridged deposit and dustcake samples taken on August 6, 2004, and on August 31, 2004. The properties of the bridged deposit most closely resemble the in situ samples collected during the use of PRB coal with dolomite addition, suggesting that the bulk of the deposit came from this part of the run. The high carbon content of the bridged deposit remaining suggests that it was not affected by the dustcake burnoff. Unlike the bridged deposit, the residual cake from August 6, 2004, had a reduced carbon content of 17.7 percent versus 31.3 percent for the in situ samples taken during the PRB with dolomite portion of the run. This result suggests that it was partially burned, but the combustion was not complete. This explains why the surface area, although reduced, was still high compared to that of combustion ash. In the past, there was an enrichment of fine particles in the residual cake. This was also evident in the cake from August 6, 2004, but some of the differences in particle size were attributed to the partial burnout.

The dustcake from August 31, 2004, was much lighter in color than the cake from August 6, 2004, suggesting that the second burnoff at the end of the run was more effective. This suggestion is confirmed by the chemical analysis, which shows the carbon content of the bulk cake was about 4.2 to 4.9 percent. The specific surface area of the cake is very low (3.5 to 3.8 m^2/g), which is another indication of very good burnout after the final shutdown. Analysis of the dustcake layers showed some variation in carbon and very little variation in the MgO content, suggesting the differences in the various layers may be primarily related to the degree of burnout. Due to the effects of the burnoff, the dustcake properties were not compared to those from previous runs.

3.2.4.4 Dustcake Flow Resistance

Lab Drag Measurements. Drag measurements were made on the two hopper samples described previously. The drag results as a function of particle size are shown in Figure 3.2-11. The data from the last run, TC15, and from previous PRB tests are also shown for comparison. Comparing only the PRB data, it is seen that the drag increases with increasing carbon content in the gasification ash. In the past, this effect has been reported with North Dakota lignites, but here the same effect was observed with PRB coal.

The hopper sample from the lignite with dolomite test did not follow the trend with carbon content, but that was not surprising since it was a different type of gasification ash with potentially different

morphological characteristics. The drag of the lignite with dolomite sample with 29 percent carbon is actually a little lower than the drag of the PRB with limestone sample with 24 percent carbon. This result may be related to the difference in surface areas ($100 \text{ m}^2/\text{g}$ for the lignite with dolomite versus $125 \text{ m}^2/\text{g}$ for the PRB with limestone). In the past, it has been seen that drag increases with increasing surface area.

Transient PCD Drag. During each in situ sampling run, the PCD transient drag was calculated using the measured pressure drop, gas flow, and particle concentration (see Table 3.2-6). The calculated transient drag at PCD conditions is listed under the column heading “PCD.” The corresponding normalized value of transient drag at room temperature is listed under the heading “PCD@RT” and is plotted as a function of non-carbonate carbon (NCC) content in Figure 3.2-12, along with data from previous PRB runs. There is considerable scatter in the data due to variations in equipment configuration, process conditions, coal composition, and limestone addition. Nevertheless, the data show a definite trend toward increasing drag with increasing carbon content. The transient drag data from the lignite + dolomite runs also show a positive correlation with carbon content, but the trend line is shifted downward from the PRB trend line. The shift could be related to the difference in surface area or to other morphological differences between the lignite + dolomite particles and the PRB + limestone particles.

Comparison of Lab Measurements with Transient Drag. It is useful to compare the lab measured drag values with the transient drag values determined from the PCD performance, because this comparison can identify situations in which the PCD pressure drop is being affected by factors other than the dustcake drag (tar deposition, failsafe plugging, etc). In the past, this comparison has been based on lab measurements made with only two or three hopper samples, which have been compared to a much larger number of PCD transient drag values that are determined on every day of sampling. The two or three hopper samples cannot accurately represent all of the individual test conditions encountered during a test condition.

In the TC15 Report, it was shown that a more reasonable comparison can be obtained by using the average lab measurements and the average PCD drag values for each given test condition (e.g., PRB + limestone and lignite + dolomite). Accordingly, the average lab and PCD drag values for each test condition are presented in Table 3.2-7. The comparison shows excellent agreement for the PRB + limestone case [49.3 versus 51.7 inwc/(lb/ft²)/(ft/min)] but worse agreement for the lignite + dolomite case [25.8 versus 41.7 inwc/(ft/min)/(lb/ft²)]. The reason for the lack of agreement in the lignite + dolomite case is unknown, but overall, the comparison between the lab and PCD drag values appears to be reasonably good.

A plot of the average lab versus PCD drag values is shown in Figure 3.2-13. This presentation shows that the data points are almost symmetrically scattered around the perfect agreement line. As illustrated in the graph, the perfect agreement line falls within the 95 percent confidence interval on the regression line to the data, indicating very good agreement between the data sets.

3.2.5 Baseline Pressure Drop Analysis

During the majority of the first portion of the test run, the baseline pressure drop was about 60 to 100 inH₂O. The baseline pressure drop increased, reaching over 130 inH₂O. This high pressure drop could have been attributed to gasification ash bridging.

During the second portion of the test run, system conditions were less stable, resulting in a less stable baseline pressure drop. While the gasifier was initially in oxygen blown mode, the baseline pressure drop was about 70 to 90 inH₂O. In air blown mode, the face velocity was much higher, and therefore the pressure drop was higher. Unsteady operation of the developmental coal feeder caused the pressure drop to fluctuate more than usual.

3.2.6 Backpulse Optimization Testing

Backpulse optimization tests were performed to monitor pressure changes at three separate locations in the PCD. Pressure taps were installed inside a filter element, inside the bottom plenum, and in the PCD vessel to evaluate the backpulse intensity. Differential pressure (DP) changes were also monitored across the failsafe tester Venturi device and across a failsafe. The backpulse pressures were varied from 150 to 300 psi above system pressure to determine the lower limits of the backpulse settings. The pressure drop response to these changes was also monitored. The pressure measurements were taken using a high speed data acquisition system. For about half of the first portion of the test run, the backpulse pressure was maintained at 150 psid above system pressure for the top and bottom plenums. Prior to TC15, the backpulse pressure was normally 400 psid for the top plenum and 600 psid for the bottom plenum. Figure 3.2-14 shows the pressure rise from the three pressure taps measured during the optimization testing. The backpulse valve open time was also varied at 0.2 and 0.4 seconds during the optimization testing. The differential pressure measurements for the 0.2 second valve open time were slightly higher than measurements for the 0.4 second time.

The backpulse frequency was also varied throughout the test run. For the first few days of the test run, the backpulse cycle time was from 10 to 20 minutes. The cycle time was increased to five minutes after resistance probes indicated gasification ash bridging. The lower backpulse pressure of 150 psi combined with longer backpulse cycle times contributed to the bridging experienced during this test run. Although the lower backpulse setting was tested briefly in TC15, it was not maintained for several days with a longer cycle time as in this test run. These less stringent settings may have had a major influence on the formation of bridging.

3.2.7 Filter Element Mechanical Testing

Pall FEAL filter element 27065 was removed from the PCD after TC16 with a cumulative total of approximately 5,060 hours of exposure in gasification operation. Mechanical testing for this element was performed at Southern Research Institute. Tests conducted included six hoop tensile tests at room temperature (RT), three axial tensile tests at RT, and three axial tensile tests at 750°F. This same test matrix has been used to test previous Pall FEAL elements after gasification operation at the PSDF. Test and specimen preparation methods were also the same as used in previous testing of FEAL elements. The axial tensile tests were conducted on dog-bone shaped specimens machined from the filter element. Hoop tensile testing was conducted using internal hydrostatic pressure on two inch long rings cut from the element with no machining of the inside diameter (ID) or outside diameter (OD) surfaces. After the element was removed from the PCD, loose gasification ash was blown off the outside surface with compressed air and no further cleaning was performed. All except for three hoop tensile specimens were tested in this condition. These three hoop tensile

specimens were ultrasonically cleaned and then de-scaled to see if that process degraded the strength of the material.

Axial tensile stress-strain curves at RT and 750°F are shown in Figure 3.2-15 and hoop tensile stress-strain curves are shown in Figure 3.2-16. Results for individual specimens and averages are given in Tables 3.2-8 and 3.2-9 for the axial and hoop directions, respectively.

Properties measured for this element, along with properties of previously tested elements, are plotted versus hours in operation in Figures 3.2-17 through 3.2-19 to show whether any degradation was evident. Axial and hoop strengths at RT are shown in Figure 3.2-17, axial strengths at RT and 750°F are shown in Figure 3.2-18, and axial strain-to-failure ratios at RT and 750°F are shown in Figure 3.2-19. The plots show no indication of degradation in strength or ductility, as indicated by the strain-to-failure, after 5,060 hrs in gasification operation. Also, the hoop strength values measured on de-scaled specimens from element 27065 were nearly the same as those measured on specimens not de-scaled. There was no indication of a strength reduction caused by the de-scaling process.

As shown in Figure 3.2-17, a previously tested Pall FEAL filter element 39151 with approximately 3,750 hours in operation that was removed after TC13 had a hoop tensile strength approximately 30 percent lower than the hoop tensile strength measured on filter elements tested earlier. The axial strength of the filter element was normal. The results for element 27065, with a similar operational history, indicate that the lower hoop tensile strength of element 39151 was a result of element-to-element variability, not degradation during gasification operation.

Table 3.2-1 PCD Inlet and Outlet Particulate Measurements

Test Date	PCD Inlet					PCD Outlet				
	Run No.	Start Time	End Time	Particle Loading,		Run No.	Start Time	End Time	H ₂ O Vapor, vol %	Particle Loading, ppmw
				ppmw	lb/hr					
Air Blown - PRB										
7/15/04	1	12:15	12:30	17900	360	1	12:10	14:40	9.6	0.40
Oxygen Blown - PRB										
7/16/04	2	13:00	13:15	30100	387	2	12:00	14:30	24.0	0.15
7/19/04	3	9:15	9:30	30600	426	3	9:00	13:00	28.0	< 0.10
7/20/04	4	9:00	9:15	34000	476 ⁽²⁾	4	8:45	12:45	28.1	< 0.10
7/21/04	5	9:30	9:45	25600	369 ⁽²⁾	5	8:30	12:00	27.7	0.11
7/22/04	6	10:35	10:50	25600	399 ⁽²⁾	6	9:30	10:00	24.9	1.59 ⁽¹⁾
7/22/04	--	--	--	--	--	7	13:30	14:00	27.0	1.15 ⁽¹⁾
7/23/04	7	9:30	9:45	19300	279	8	10:10	11:10	26.7	3.63 ⁽¹⁾
7/26/04	8	8:35	8:50	18800	290 ⁽³⁾	9	8:30	12:30	28.5	< 0.10
7/27/04	9	9:50	10:05	23600	361 ⁽³⁾	10	9:30	13:30	27.9	< 0.10
7/28/04	10	10:10	10:25	28400	450 ⁽³⁾	11	10:00	14:00	27.5	< 0.10
7/29/04	11	9:00	9:15	14300	226 ⁽³⁾	--	--	--	--	--
Air Blown - PRB										
7/30/04	12	12:30	12:45	9600	207	12	11:00	14:00	17.3	< 0.10
8/2/04	13	10:35	10:50	13100	284	13	10:30	14:30	9.8	< 0.10
8/3/04	--	--	--	--	--	14	11:45	14:45	11.4	< 0.10
Oxygen Blown (High Pressure) - PRB										
8/11/04	14	11:00	11:15	18000 ⁽³⁾	286 ⁽³⁾	15	8:45	14:45	26.5	(4)
8/12/04	15	9:25	9:40	15400 ⁽³⁾	260 ⁽³⁾	16	9:15	13:15	29.2	0.11
8/13/04	16	12:45	13:00	21000 ⁽³⁾	365 ⁽³⁾	17	10:30	14:30	28.5	< 0.10
Air Blown - PRB										
8/16/04	17	9:15	9:30	14400 ⁽³⁾	334 ⁽³⁾	18	9:00	13:00	11.3	< 0.10
Air Blown - Lignite										
8/17/04	18	13:15	13:30	36000 ⁽³⁾	621 ⁽³⁾	19	10:00	14:00	19.2	< 0.10
8/18/04	19	12:30	12:45	34700 ⁽³⁾	595 ⁽³⁾	20	10:15	14:15	21.0	< 0.10
8/19/04	20	10:00	10:15	44300 ⁽³⁾	777 ⁽³⁾	21	9:45	13:45	18.8	< 0.10
8/20/04	21	9:15	9:30	35300 ⁽³⁾	585 ⁽³⁾	22	9:00	13:00	18.2	< 0.10
8/23/04	22	9:05	9:20	38100 ⁽³⁾	691 ⁽³⁾	23	9:00	13:00	15.3	< 0.10
8/24/04	23	9:15	9:30	34700 ⁽³⁾	633 ⁽³⁾	24	9:00	13:00	15.2	< 0.10
Notes: 1. Dust injection for PCME testing. 2. Limestone addition to gasifier. 3. Dolomite addition to gasifier. 4. Tar on filter and blank precludes accurate mass. Some increased particle loading. Sample was paused from 8:50 until 10:45 because of oxygen trip.										

Table 3.2-2 Physical Properties In Situ Samples and Hopper Samples

Sample ID	Run No.	Sample Date	Bulk Density g/cc	True Density g/cc	Uncompacted Bulk Porosity %	BET Specific Surface Area m ² /g	Mass Median Diameter μm
PRB Coal - No Sorbent							
AB15508	1	7/15/04	0.28	2.30	87.8	194	17.6
AB15509	2	7/16/04	0.32	2.44	86.9	136	25.8
AB15510	3	7/19/04	0.32	2.38	86.6	162	18.9
<i>Average</i>			0.31	2.37	87.1	164	20.7
PRB Coal with Limestone							
AB15511	4	7/20/04	0.31	2.43	87.2	161	21.3
AB15512	5	7/21/04	0.35	2.38	85.3	124	23.0
AB15513	6	7/22/04	0.31	2.52	87.7	111	15.5
<i>Average</i>			0.32	2.44	86.7	132	19.9
PRB Coal - No Sorbent							
AB15514	7	7/23/04	0.23	2.39	90.4	173	15.5
PRB Coal with Dolomite							
AB15515	8	7/26/04	0.23	2.28	89.9	245	13.4
AB15516	9	7/27/04	0.26	2.34	88.9	202	14.1
AB15517	10	7/28/04	0.28	2.31	87.9	214	19.2
AB15518	11	7/29/04	0.20	2.66	92.5	162	12.4
AB15519	12	7/30/04	0.22	2.78	92.1	173	13.8
AB15520	13	8/2/04	0.21	2.55	91.8	289	18.3
AB15849	14	8/11/04	0.25	2.39	89.5	136	16.5
AB15850	15	8/12/04	0.25	2.21	88.7	126	14.2
AB15851	16	8/13/04	0.21	2.45	91.4	189	14.1
AB15852	17	8/16/04	0.27	2.43	88.9	233	16.9
<i>Average</i>			0.24	2.44	90.2	197	15.3
Lignite with Dolomite							
AB15853	18	8/17/04	0.46	2.29	79.9	51	17.2
AB15854	19	8/18/04	0.55	2.20	75.0	53	21.7
AB15855	20	8/19/04	0.52	2.36	78.0	58	18.6
AB15856	21	8/20/04	0.62	2.55	75.7	40	13.6
AB15857	22	8/23/04	0.49	2.38	79.4	119	12.6
AB15858	23	8/24/04	0.46	2.38	80.7	100	17.3
<i>Average</i>			0.52	2.36	78.1	70	16.8
Hopper Samples							
AB15533	1	7/21/04	0.36	2.50	85.6	125	16.9
AB15881	2	8/24/04	0.49	2.34	79.1	100	12.9

Table 3.2-3 Chemical Composition of In Situ Samples and Composite Hopper Samples

Sample ID	Run No.	Sample Date	CaCO ₃ Wt %	CaS Wt %	CaO Wt %	MgO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %	Loss on Ignition Wt %	Molar Ratio Ca/Mg
PRB Coal - No Sorbent										
AB15508	1	07/15/04	8.89	1.13	4.38	2.51	28.80	54.29	31.56	2.91
AB15509	2	07/16/04	10.20	0.16	4.99	2.17	19.83	62.65	26.10	3.57
AB15510	3	07/19/04	9.84	0.07	4.93	2.44	24.36	58.36	26.69	3.07
<i>Average</i>			9.64	0.45	4.77	2.37	24.33	58.43	28.12	3.19
PRB Coal with Limestone										
AB15511	4	07/20/04	10.68	0.07	15.42	5.20	26.46	42.17	29.70	2.94
AB15512	5	07/21/04	10.41	0.07	13.84	4.59	25.91	45.18	30.40	3.07
AB15513	6	07/22/04	10.23	0.04	21.60	6.93	21.03	40.17	25.27	2.82
<i>Average</i>			10.44	0.06	16.95	5.57	24.47	42.51	28.46	2.95
PRB Coal - No Sorbent										
AB15514	7	07/23/04	9.39	0.09	12.37	3.84	33.60	40.71	37.52	3.29
PRB Coal with Dolomite										
AB15515	8	07/26/04	9.95	0.07	12.09	7.64	40.31	29.94	44.73	1.66
AB15516	9	07/27/04	9.61	0.07	14.34	9.41	35.01	31.56	40.08	1.50
AB15517	10	07/28/04	10.59	0.09	10.80	7.84	43.00	27.69	48.63	1.53
AB15518	11	07/29/04	9.27	0.04	18.87	11.37	24.68	35.76	29.01	1.51
AB15519	12	07/30/04	9.55	0.07	16.37	7.48	22.52	44.01	27.22	2.08
AB15520	13	08/02/04	6.93	1.02	11.88	4.06	36.11	40.00	37.46	2.91
AB15849	14	08/11/04	13.70	0.07	9.64	5.83	28.73	42.03	33.67	2.13
AB15850	15	08/12/04	14.00	0.02	13.03	8.25	24.60	40.10	28.61	1.81
AB15851	16	08/13/04	24.68	0.07	5.83	8.63	31.60	29.20	38.33	1.63
AB15852	17	08/16/04	24.64	0.13	8.02	11.30	26.52	29.39	32.70	1.38
<i>Average</i>			13.29	0.16	12.09	8.18	31.31	34.97	36.04	1.81
Lignite with Dolomite										
AB15853	18	08/17/04	31.30	0.20	1.01	9.85	30.80	26.85	42.38	1.36
AB15854	19	08/18/04	29.23	0.09	4.01	11.23	23.82	31.62	36.33	1.30
AB15855	20	08/19/04	31.64	0.09	1.98	11.36	26.92	28.01	39.62	1.24
AB15856	21	08/20/04	32.25	0.02	6.45	15.54	18.48	27.25	33.05	1.13
AB15857	22	08/23/04	32.61	0.16	0.00	10.63	31.38	25.48	43.41	1.23
AB15858	23	08/24/04	30.98	0.11	2.84	12.46	27.29	26.32	38.35	1.16
<i>Average</i>			31.33	0.11	2.72	11.85	26.45	27.59	38.86	1.24
Hopper Samples										
AB15533	1	07/21/04	7.55	0.04	18.05	5.32	23.97	45.07	26.36	3.00
AB15881	2	08/24/04	23.66	0.09	6.30	11.85	29.41	28.69	39.80	1.18

Table 3.2-4 Physical Properties of Bridged Deposit and Dustcake Samples

Sample ID	Sample Date	Bulk Density g/cc	True Density g/cc	Uncompacted Bulk Porosity %	Specific Surface Area m ² /g	Mass-Median Diameter μm	Loss on Ignition Wt %
<i>Bridged Deposit Sampled through Manway</i>							
AB15534	08/06/04	0.23	2.48	90.7	199	15.6	37.60
<i>Residual Dustcake Sampled through Manway</i>							
AB15535	08/06/04	0.38	2.77	86.3	102	10.1	20.12
<i>Outer, Middle, and Inner Dustcake Layers from Bottom Plenum (After Burnoff)</i>							
AB15912	08/31/04	0.65	2.80	76.8	2.6	8.2	39.12
AB15913	08/31/04	0.85	2.84	70.1	4.4	30.4	23.57
AB15914	08/31/04	0.72	2.86	74.8	6.6	10.5	27.57
<i>Outer and Middle+Inner Dustcake Layers from Top Plenum (After Burnoff)</i>							
AB15915	08/31/04	0.58	2.84	79.6	2.9	5.6	41.78
AB15916	08/31/04	0.74	2.90	74.5	4.2	6.6	29.47
<i>Bulk Dustcake Samples from Top and Bottom Plenums</i>							
AB15917	08/31/04	0.78	2.96	73.6	3.8	8.9	32.20
AB15918	08/31/04	0.81	2.88	71.9	3.5	13.3	
<i>Average</i>		0.80	2.92	72.8	3.6	11.1	32.20

Table 3.2-5 Chemical Composition of Bridged Deposit and Dustcake Samples

Sample ID	Sample Date	CaCO ₃ Wt %	CaS Wt %	CaO Wt %	MgO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %	Loss on Ignition Wt %	Molar Ratio Ca/Mg
<i>Bridged Deposit Sampled Through Manway</i>									
AB15534	08/06/04	7.02	0.56	10.63	3.66	33.75	44.38	37.60	2.93
<i>Residual Dustcake Sampled Through Manway</i>									
AB15535	08/06/04	6.77	0.80	16.08	5.39	17.66	53.29	20.12	2.72
<i>Outer, Middle, and Inner Dustcake Layers from Bottom Plenum (After Burnoff)</i>									
AB15912	08/31/04	51.27	0.27	0.00	19.99	6.45	22.03	39.12	1.03
AB15913	08/31/04	49.61	0.63	0.00	18.91	1.87	28.99	23.57	1.07
AB15914	08/31/04	49.44	0.56	0.00	18.94	3.39	27.68	27.57	1.06
<i>Outer and Middle+ Inner Dustcake Layers from Top Plenum (After Burnoff)</i>									
AB15915	08/31/04	51.68	0.20	0.00	20.22	6.95	20.95	41.78	1.03
AB15916	08/31/04	51.23	0.42	0.00	20.30	3.82	24.22	29.47	1.02
<i>Bulk Dustcake Samples from Top and Bottom Plenums (After Burnoff)</i>									
AB15917	08/31/04	50.25	0.60	0.00	19.62	4.93	24.60	32.20	1.04
AB15918	08/31/04	51.12	0.49	0.00	19.36	4.23	24.81	31.53	1.07
<i>Average</i>		50.69	0.55	0.00	19.49	4.58	24.70	31.87	1.06

Table 3.2-6 Transient Drag Determined from PCD ΔP and from Lab Measurements

Run No.	$\Delta P/\Delta t$, inwc/min	$\Delta(AL)/\Delta t$, lb/ft ² /min	FV, ft/min	MMD, μm	NCC, %	Drag, inwc/(lb/ft ²)/(ft/min)		
						PCD	PCD@RT	Lab
PRB + Limestone								
4	2.48	0.037	3.12	21.3	26.5	67	43	50
5	2.43	0.028	3.15	23.0	25.9	86	56	40
6	2.39	0.031	3.48	15.5	21.0	78	49	65
AVG	2.43	0.03	3.25	19.9	24.5	77	49	52
Lignite + Dolomite								
18	2.53	0.048	3.88	17.2	30.8	53	33	38
19	1.69	0.046	3.83	21.7	23.8	37	24	31
20	2.33	0.060	3.91	18.6	26.9	39	25	35
21	1.24	0.045	3.64	13.6	18.5	27	18	53
22	2.32	0.053	4.01	12.6	31.4	44	28	55
23	2.11	0.049	4.06	17.3	27.3	43	27	38
AVG	2.04	0.050	3.89	16.8	26.5	41	26	42
Note: Lab drag data calculated from linear regression to MMD.								

Nomenclature:

$\Delta P/\Delta t$ = rate of pressure drop rise during particulate sampling run, inwc/min.

$\Delta(AL)/\Delta t$ = rate of increase in areal loading during sampling run, lb/min/ft².

FV = average PCD face velocity during particulate sampling run, ft/min.

MMD = mass-median diameter of in situ particulate sample, μm .

NCC = non-carbonate carbon.

RT = room temperature, 77°F (25°C).

Table 3.2-7 Comparison of Average Drag Values Determined from PCD Performance and from Lab Measurements

Run	Coal	Average Transient Drag Determined from PCD Performance, inwc/(lb/ft ²)/(ft/min)	Average Drag Determined from RAPTOR Lab Measurements, inwc/(lb/ft ²)/(ft/min)	Difference, %
GCT2	PRB	29.3	20.9	-33.5
GCT3	PRB	80.2	92.7	14.5
GCT4	PRB	66.4	57	-15.2
TC06	PRB	89.4	81.2	-9.6
TC07	PRB	47.7	49.8	4.3
TC08	PRB	46.5	50	7.3
TC09	Hiawatha Bit	29.0	23.3	-21.8
TC10	PRB	44.7	57.6	25.2
TC11	Falkirk Lignite	16.1	35.9	76.2
TC12	PRB	58.0	60.8	4.7
TC13	Freedom Lignite	34.4	39.4	13.6
TC14	PRB	47.4	41.6	-13.0
TC15	PRB	54.6	76.4	33.3
TC16	PRB + Limestone	49.3	51.7	4.8
TC16	Lignite + Dolomite	25.8	41.7	47.1
<i>Average</i>		47.9	52.0	9.2

Table 3.2-8 Axial Tensile Results for Pall FEAL Filter Element 27065

Specimen Number	Hours in Operation	Test Temperature (°F)	Ultimate Strength (psi)	Young's Modulus (Msi)	Strain-to-Failure (mils/in)
Tn-Ax-118	5057	RT	16.2	5.47	4.8
Tn-Ax-120	5057	RT	18.0	5.43	6.5
Tn-Ax-122	5057	RT	17.1	5.91	5.5
Average			17.1	5.60	5.6
Tn-Ax-119	5057	750	18.4	5.02	16.2
Tn-Ax-121	5057	750	19.6	5.02	17.8
Tn-Ax-123	5057	750	17.8	4.81	14.9
Average			18.6	4.95	16.3

Table 3.2-9 Hoop Tensile Results for Pall FEAL Filter Element 27065

Specimen Number	Hours in Operation	Maximum Hydrostatic Pressure (psig)	Ultimate Strength (Ksi)	Young's Modulus (Msi)	Maximum Strain at OD (mils/in)	Remarks
Tn-Hoop-1A	5057	980	13.6	4.97	3.36	
Tn-Hoop-1B	5057	870	12.8	5.82	2.42	
Tn-Hoop-1C	5057	1020	14.6	5.73	3.15	Descaled
Tn-Hoop-2A	5057	910	13.2	5.19	2.97	Descaled
Tn-Hoop-2B	5057	680	12.2	5.61	2.3	
Tn-Hoop-2C	5057	1010	14.3	5.56	3.07	Descaled
Average		912	13.5	5.48	2.88	
Standard Deviation		117	0.82	0.3	0.39	
Coefficient of Variation		13%	6.1%	5.5%	13%	

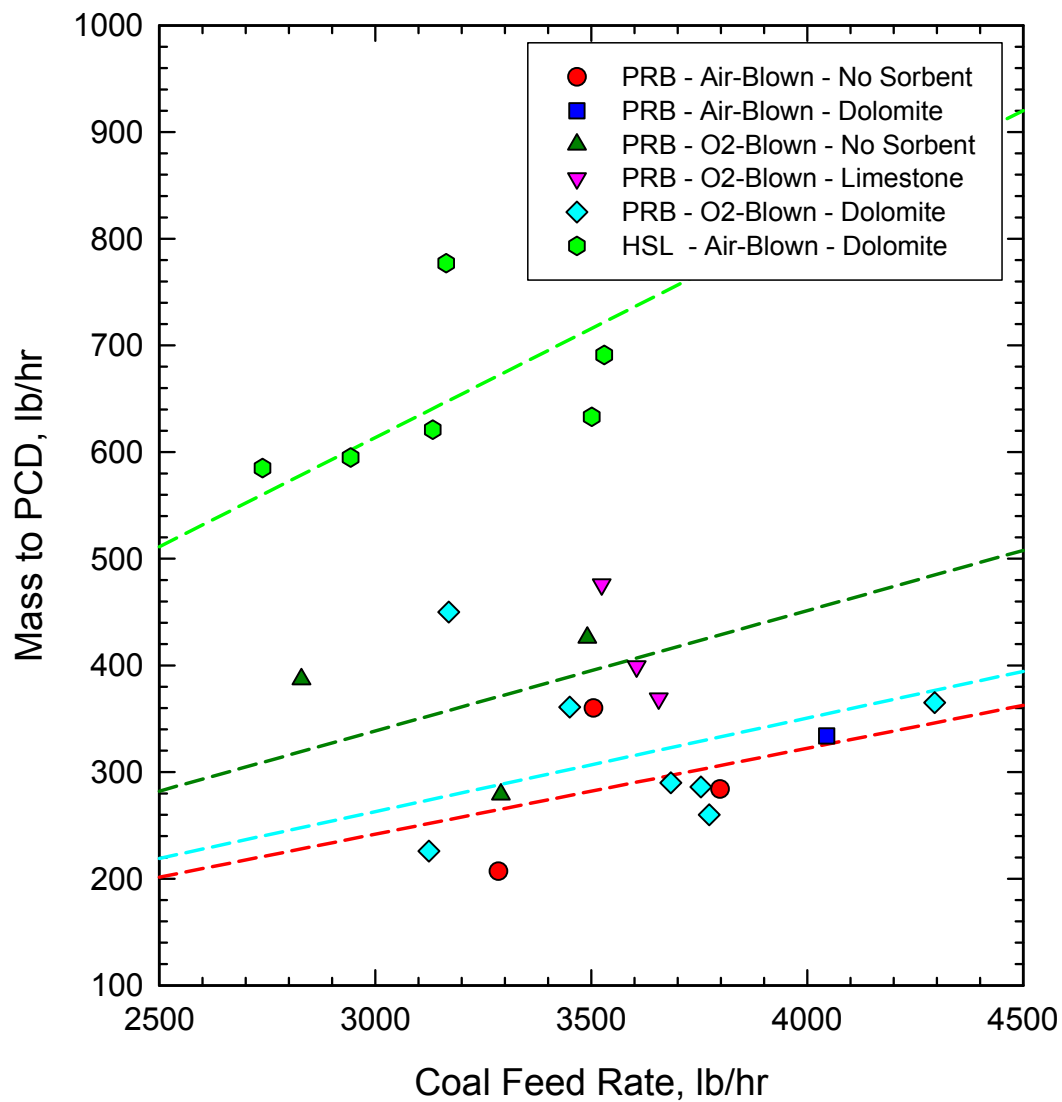


Figure 3.2-1 PCD Inlet Particle Concentration as a Function of Coal Feed Rate.

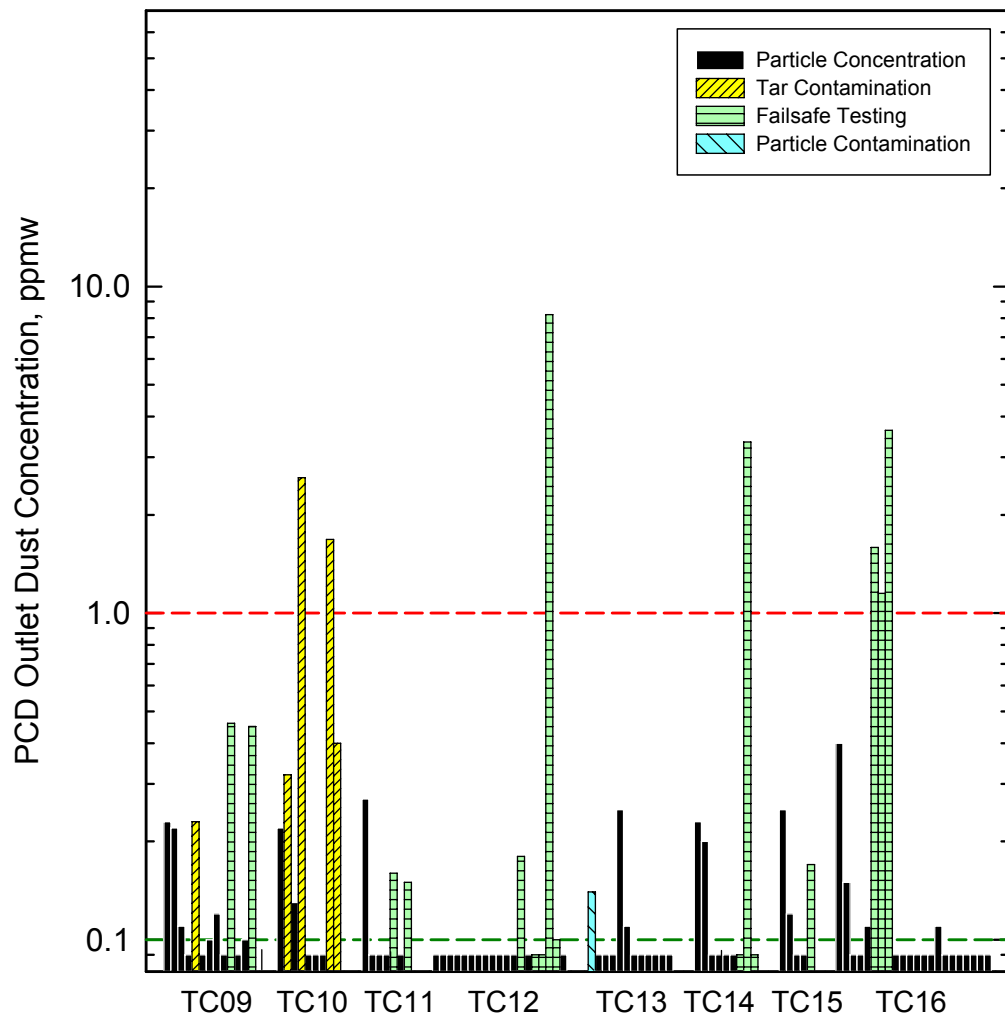


Figure 3.2-2 PCD Outlet Dust Concentration for Recent Gasification Runs



Figure 3.2-3 Gasket Removed and Dissected After TC16

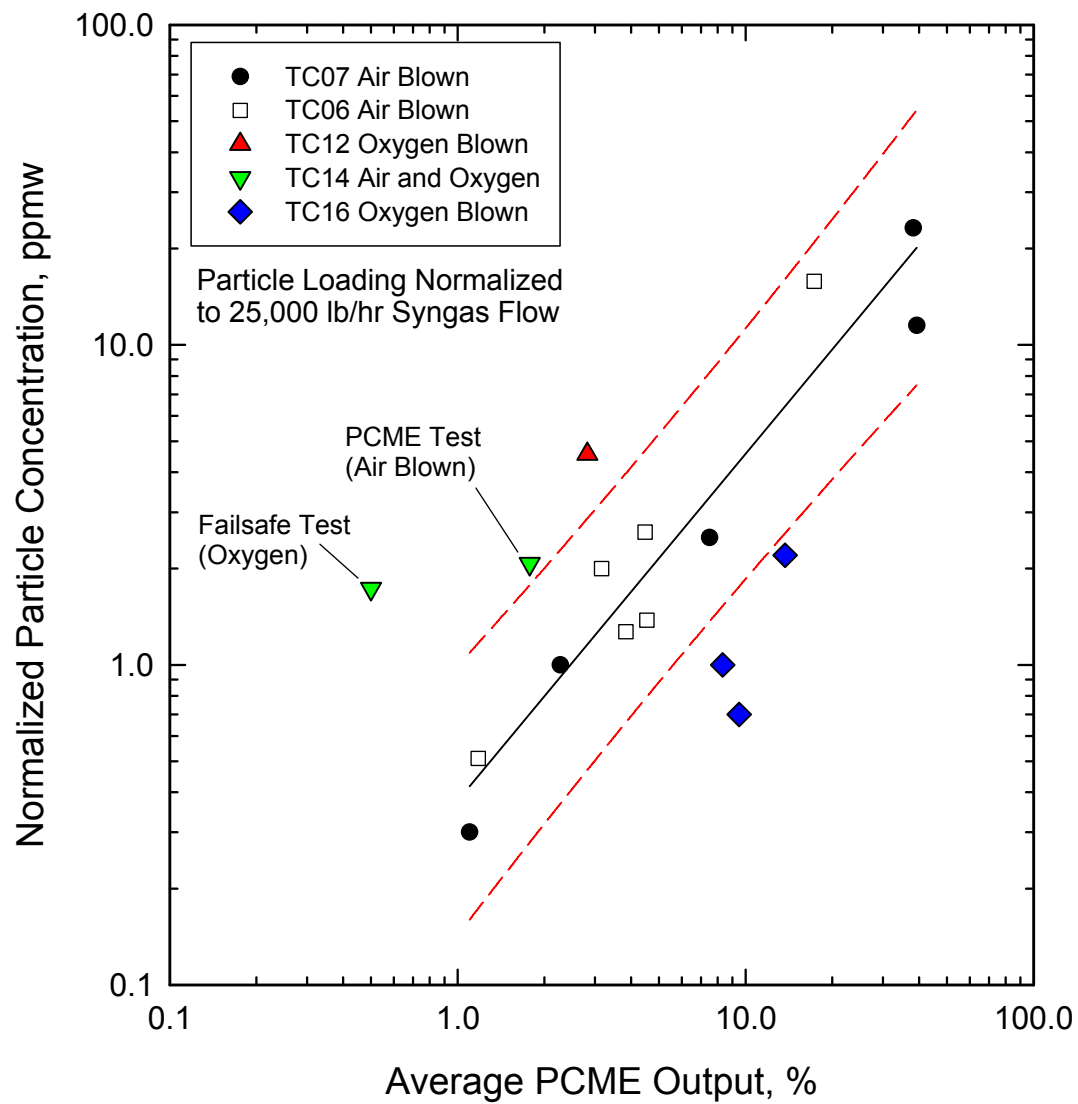
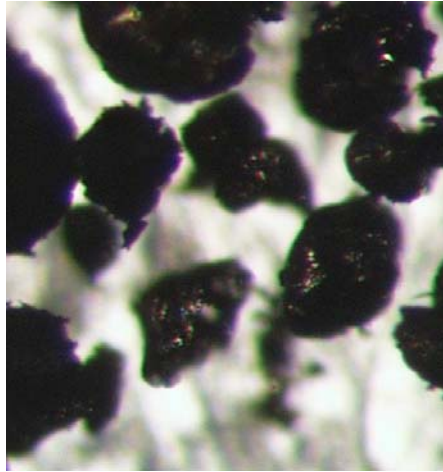
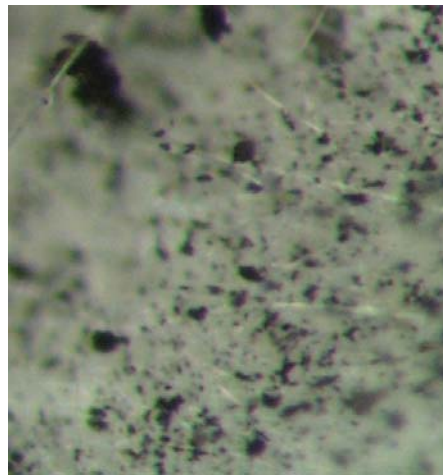


Figure 3.2-4 Response of PCME Particulate Monitor in TC16 and Earlier Test Runs



TC16 PCD Outlet Sample No. 6 - Large Particles Dumped Off Filter
Measured Loading = 1 ppmw* - PCME Response = 8.3 %



TC14 PCD Outlet Sample No. 7 - Particles on Sample Filter
Measured Loading = 2 ppmw* - PCME Response = 1.8 %

Figure 3.2-5 Comparison of Large Particle Fraction in TC16 and TC14 PCME Injection Tests

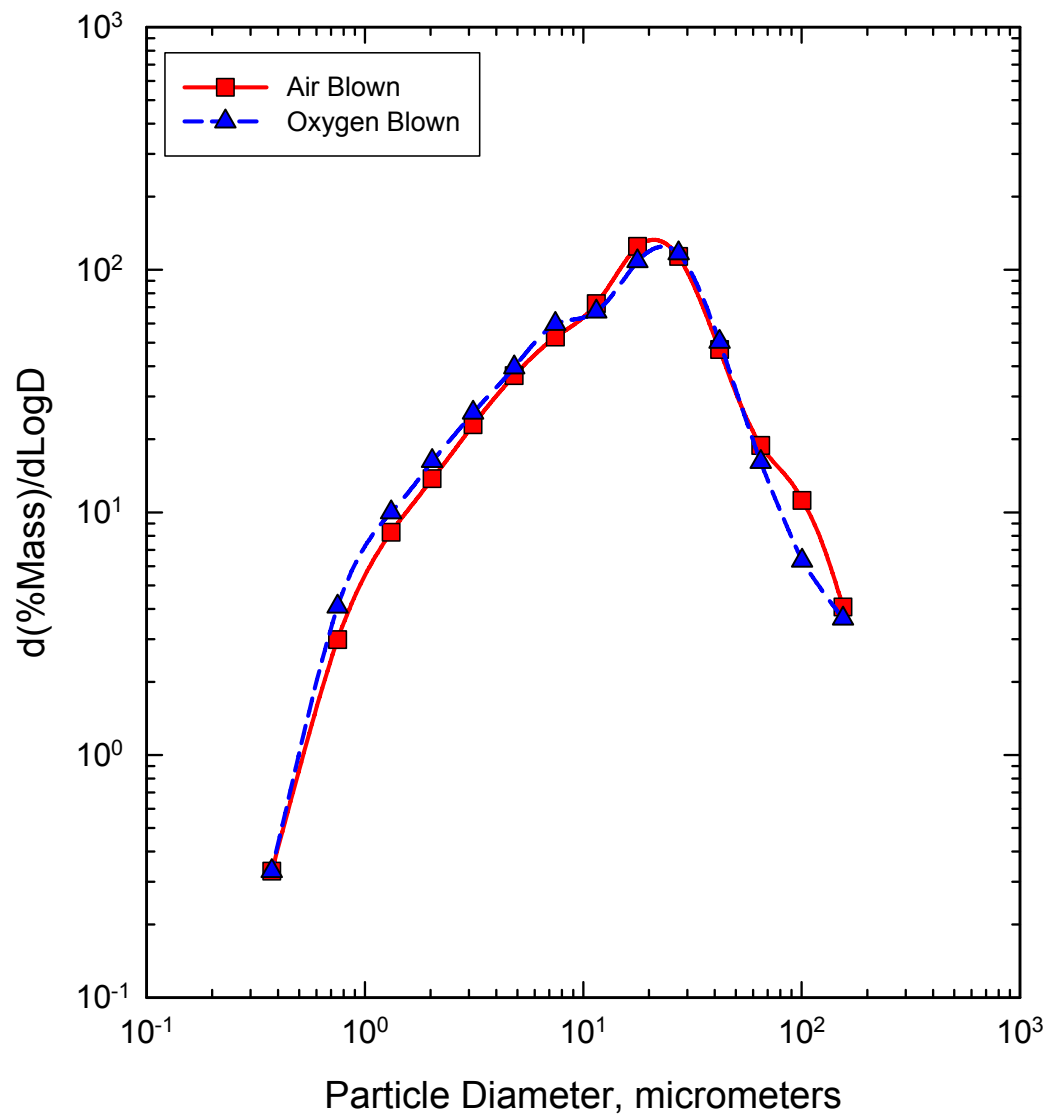


Figure 3.2-6 Effect of Gasification Oxidant on PCD Inlet Particle Size Distribution with PRB Coal

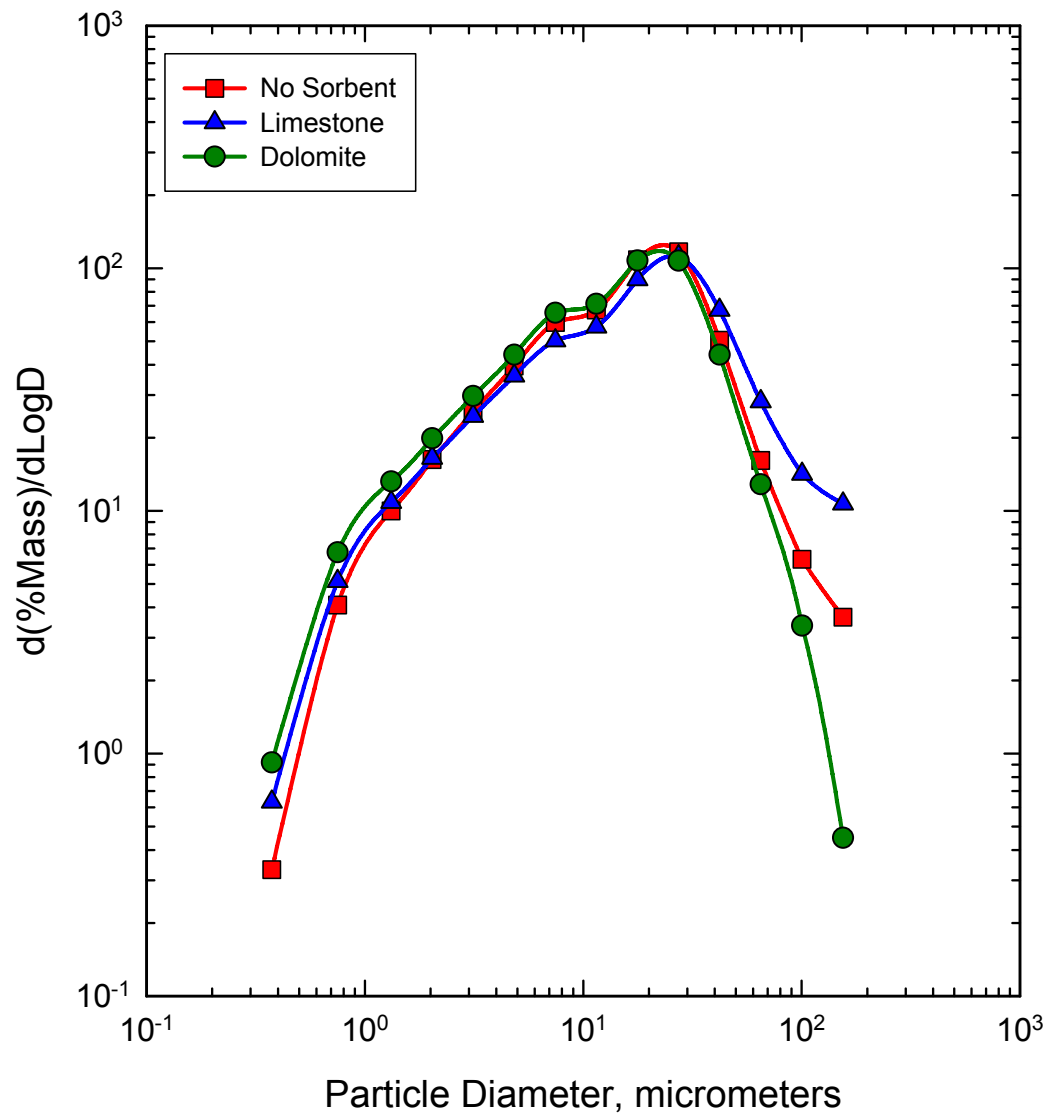


Figure 3.2-7 Effect of Sorbent on PCD Inlet Particle Size Distribution with PRB Coal

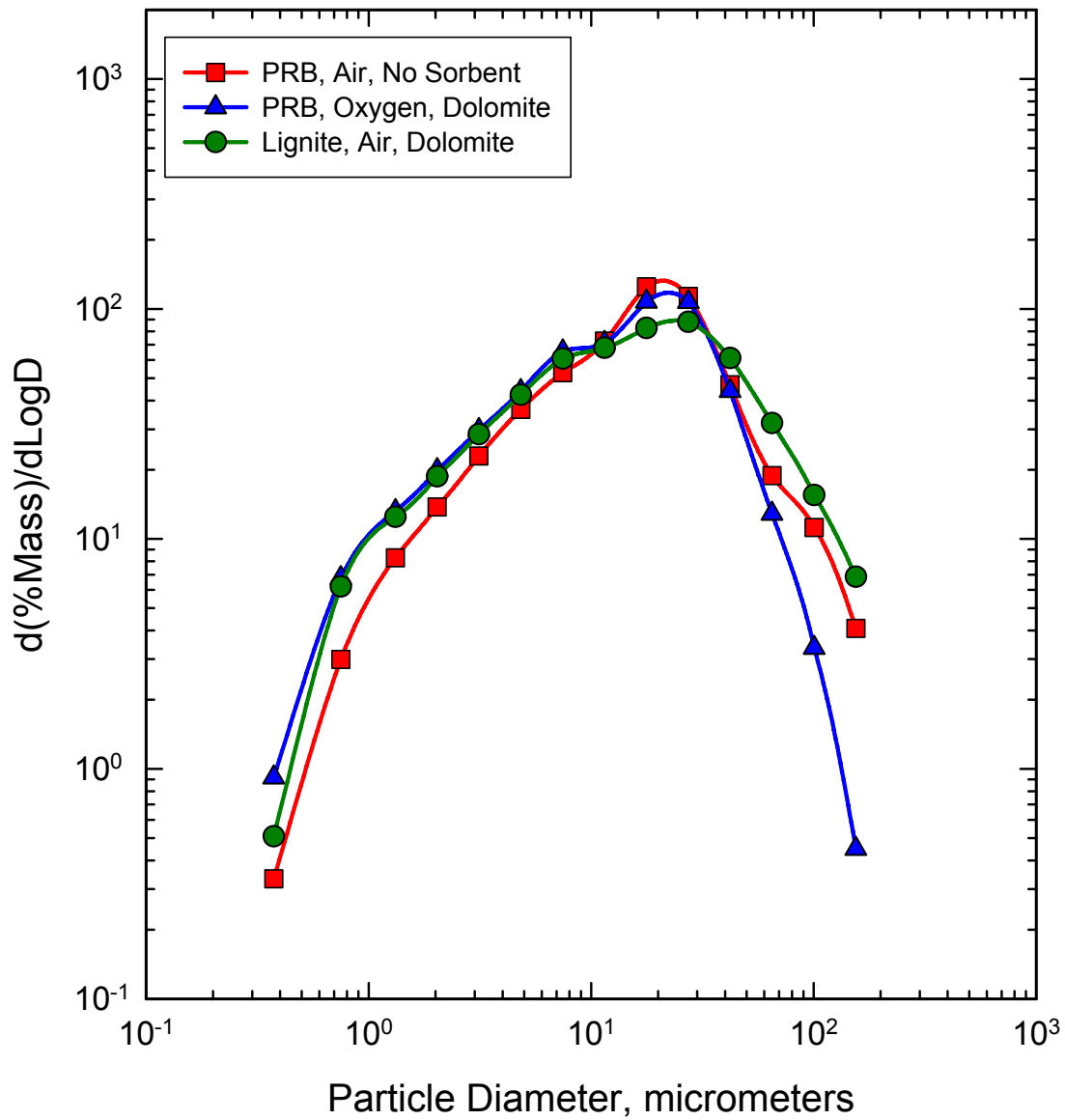


Figure 3.2-8. Effect of Coal Type on PCD Inlet Particle Size Distribution

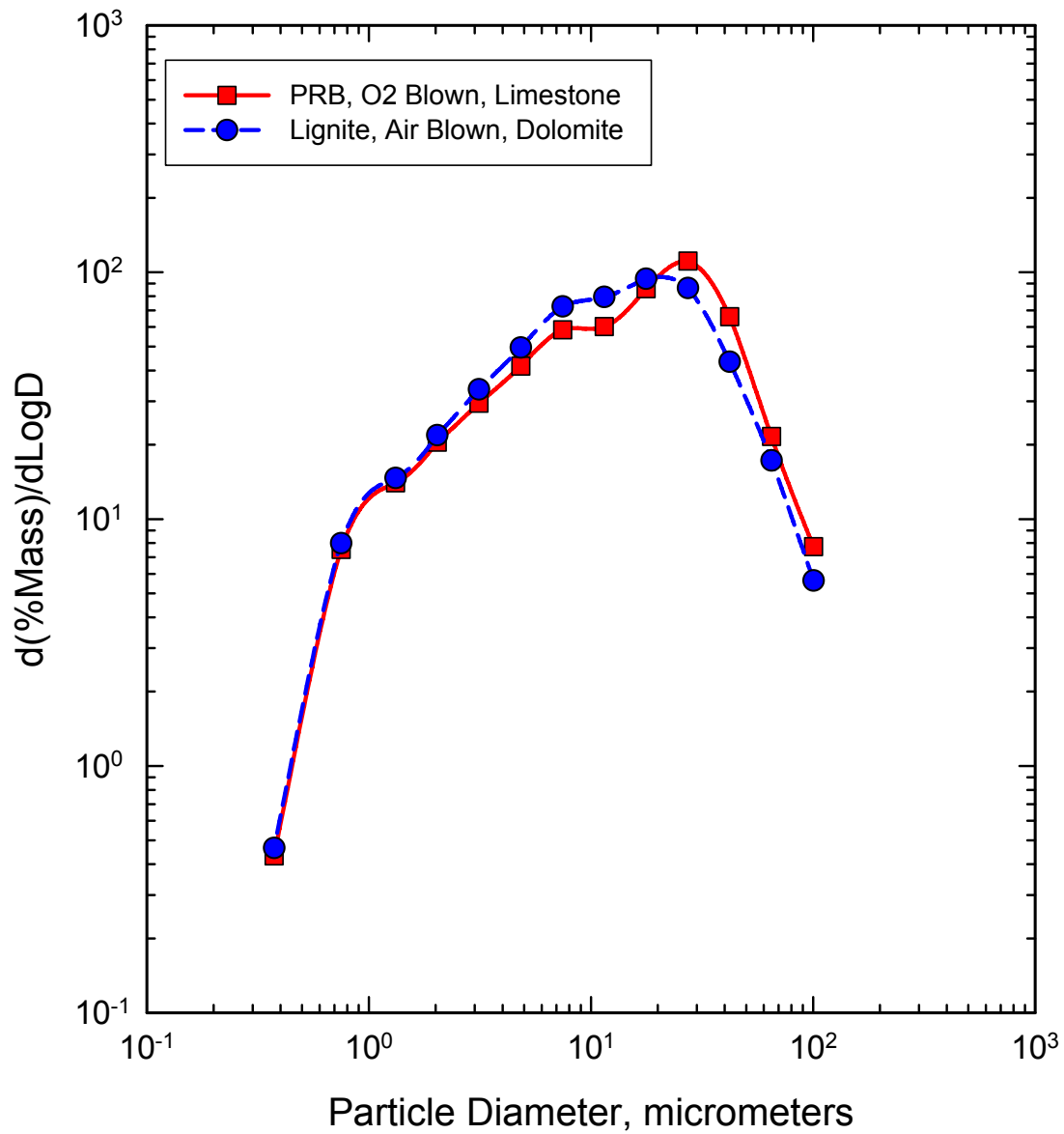


Figure 3.2-9. Particle Size Distribution of PCD Hopper Samples

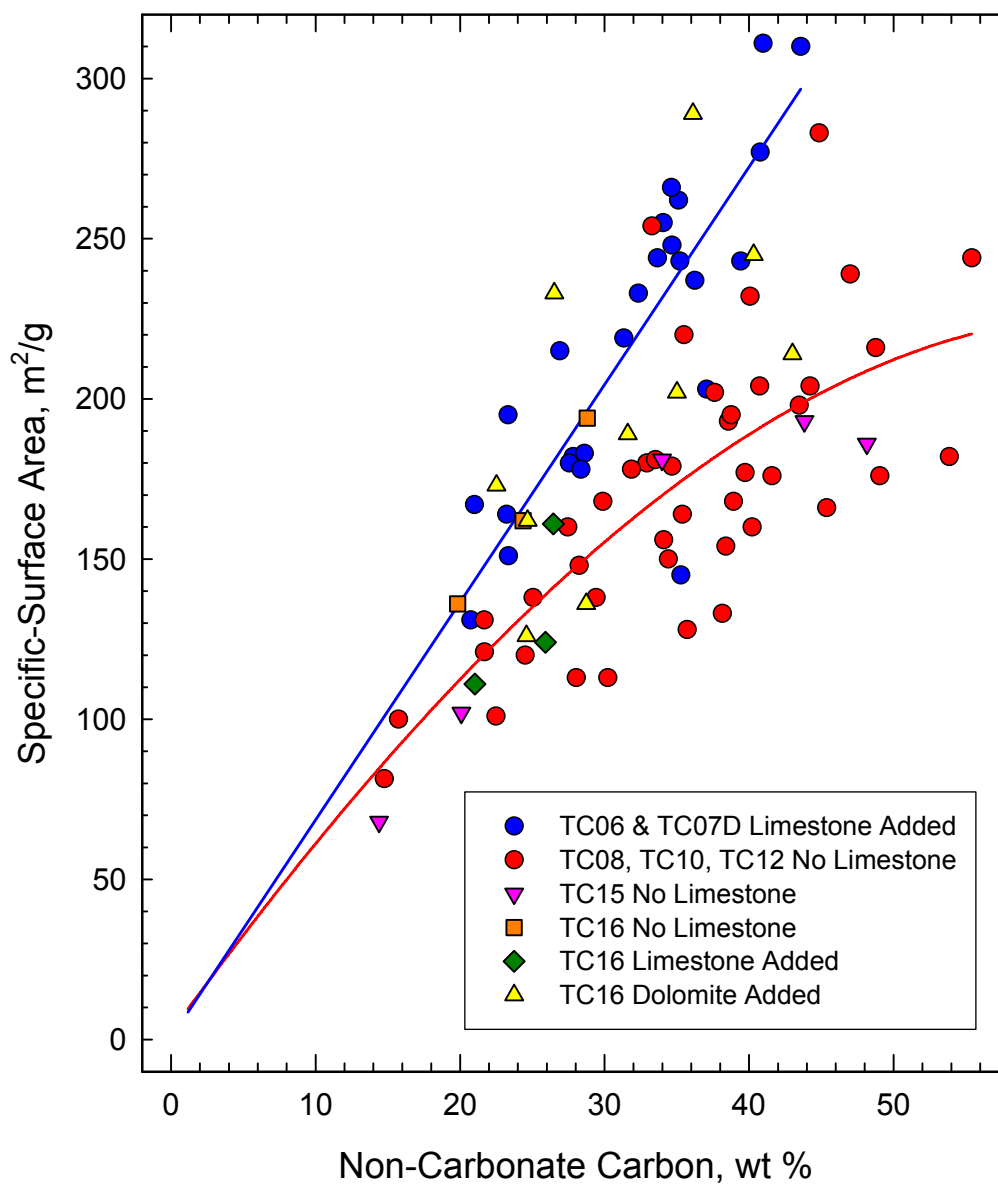


Figure 3.2-10 Specific Surface Area versus Carbon Content of In Situ Samples

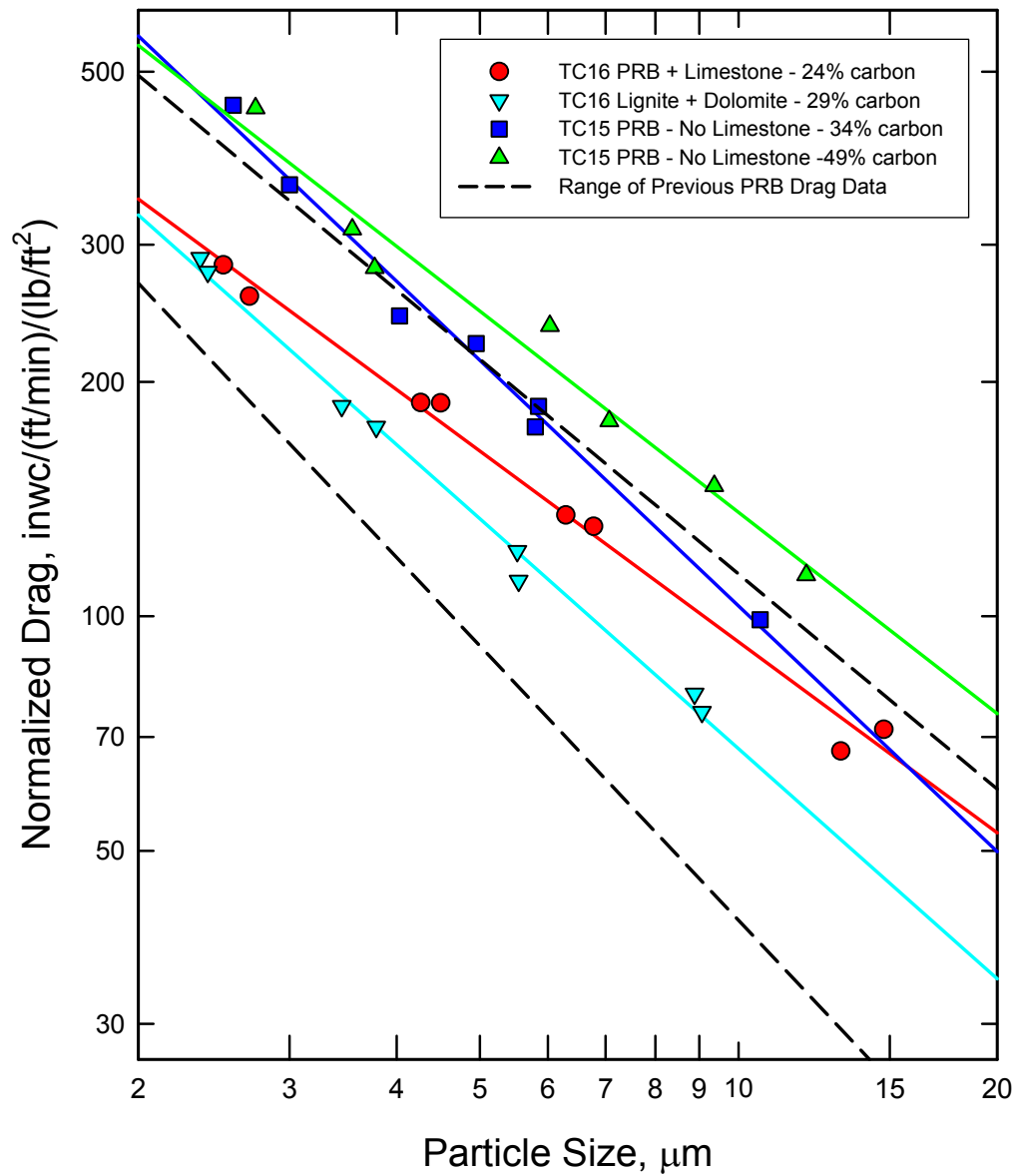


Figure 3.2-11 Lab Measured Drag as a Function of Particle Size

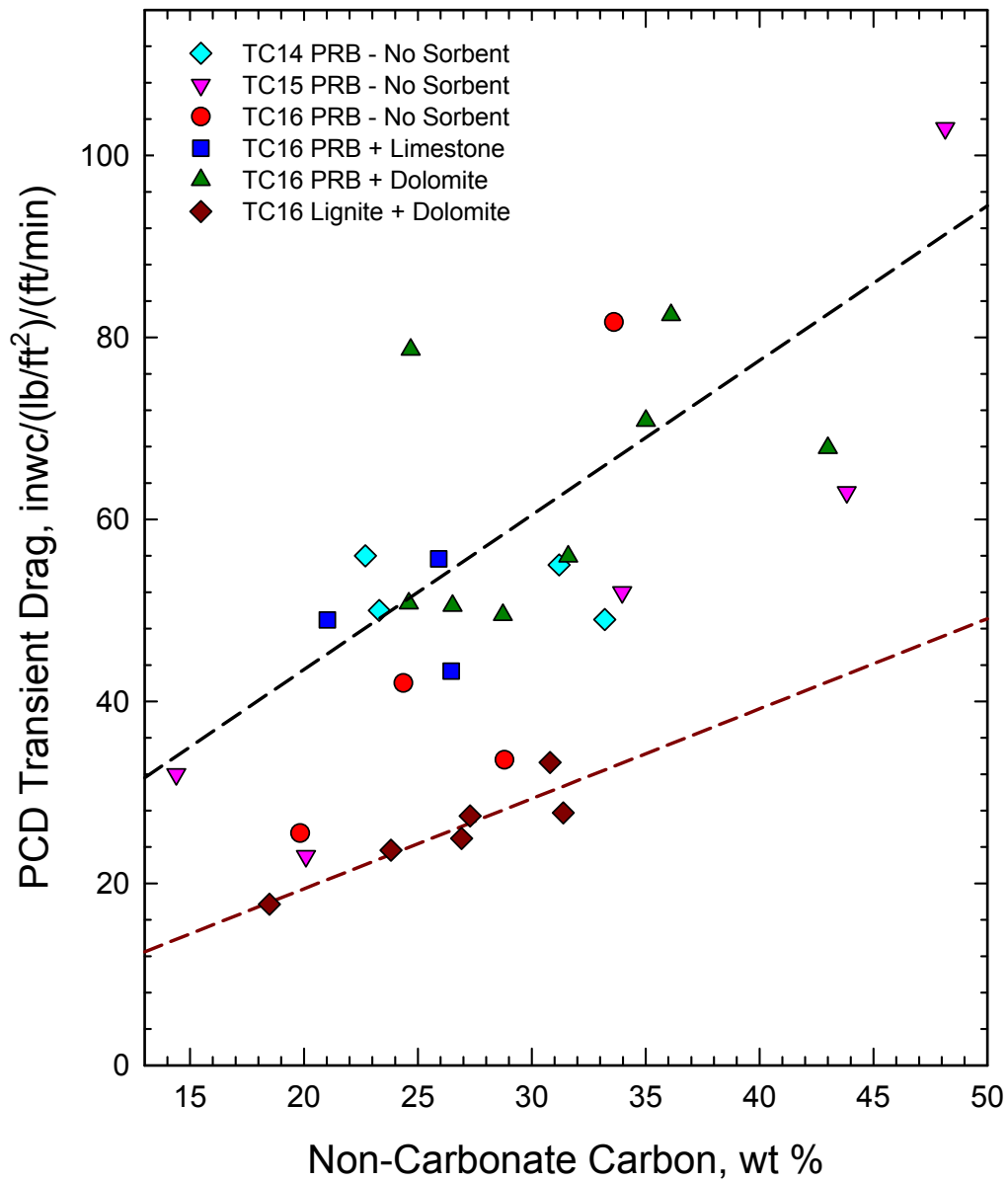


Figure 3.2-12 PCD Transient Drag versus Carbon Content of In Situ Samples.

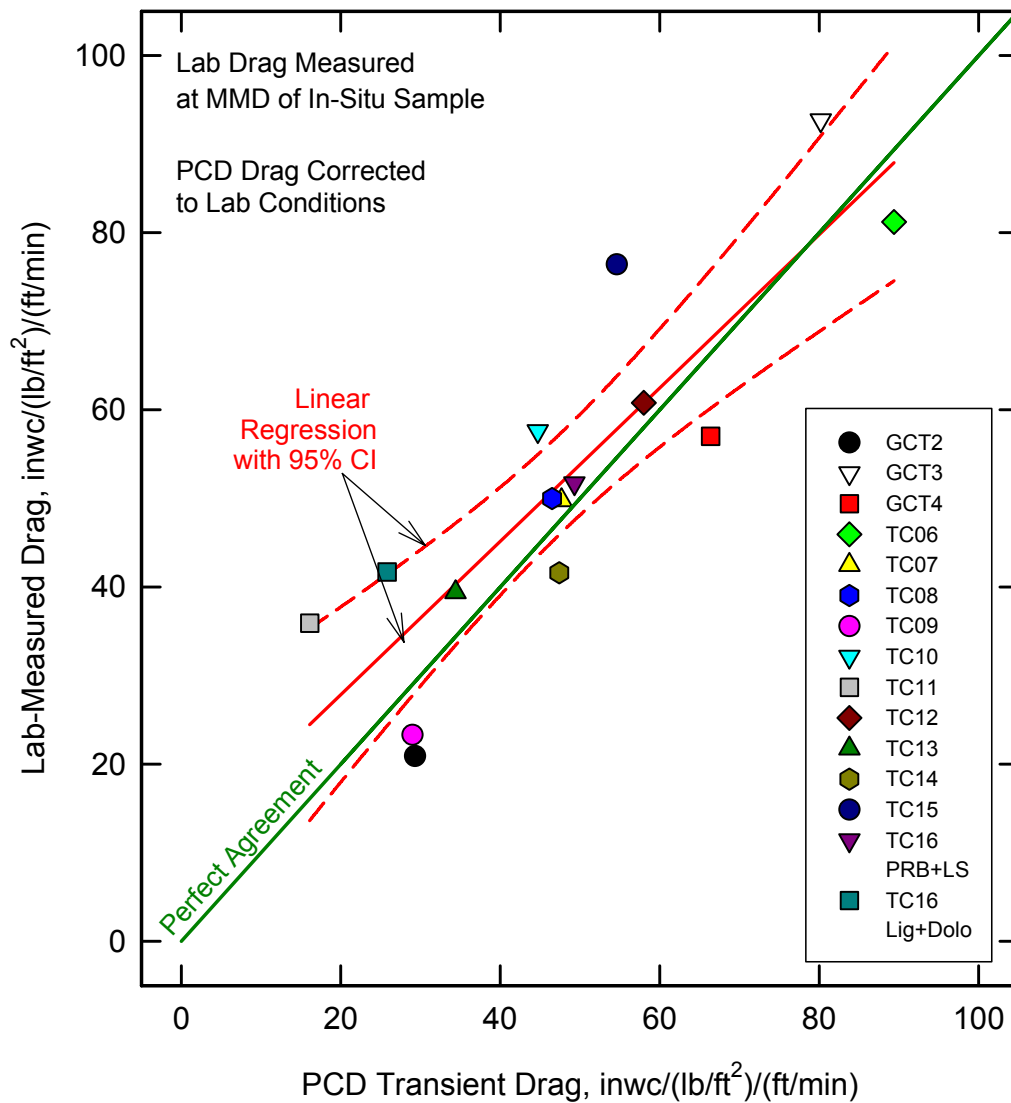


Figure 3.2-13 Comparison of PCD Transient Drag with Laboratory Measurements

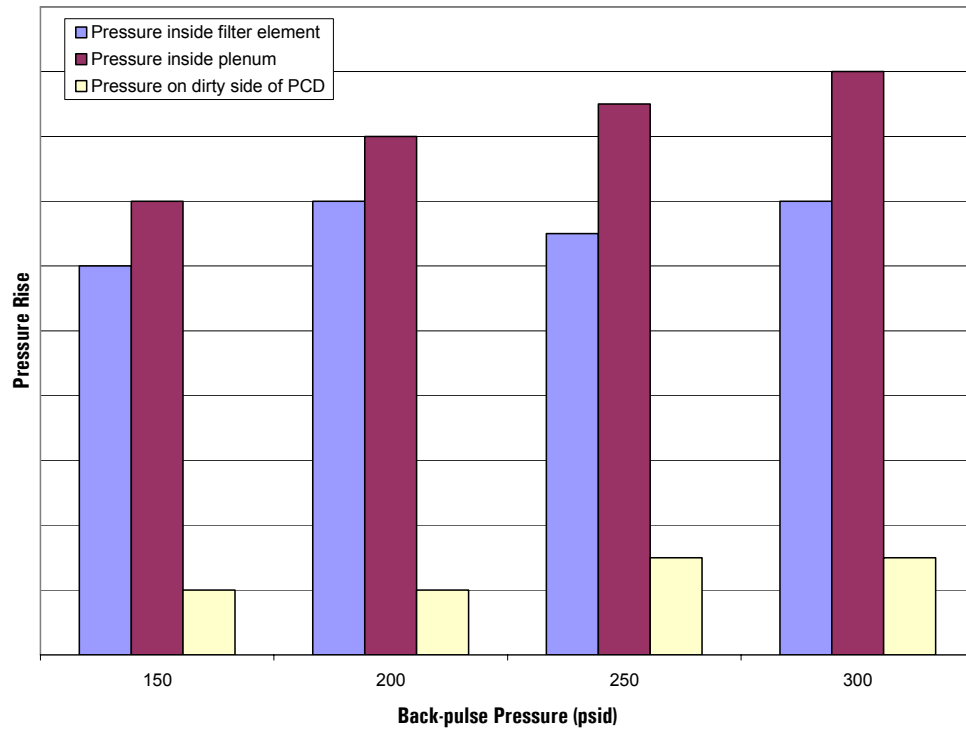


Figure 3.2-14 Pressure Rise during Backpulse Optimization Testing

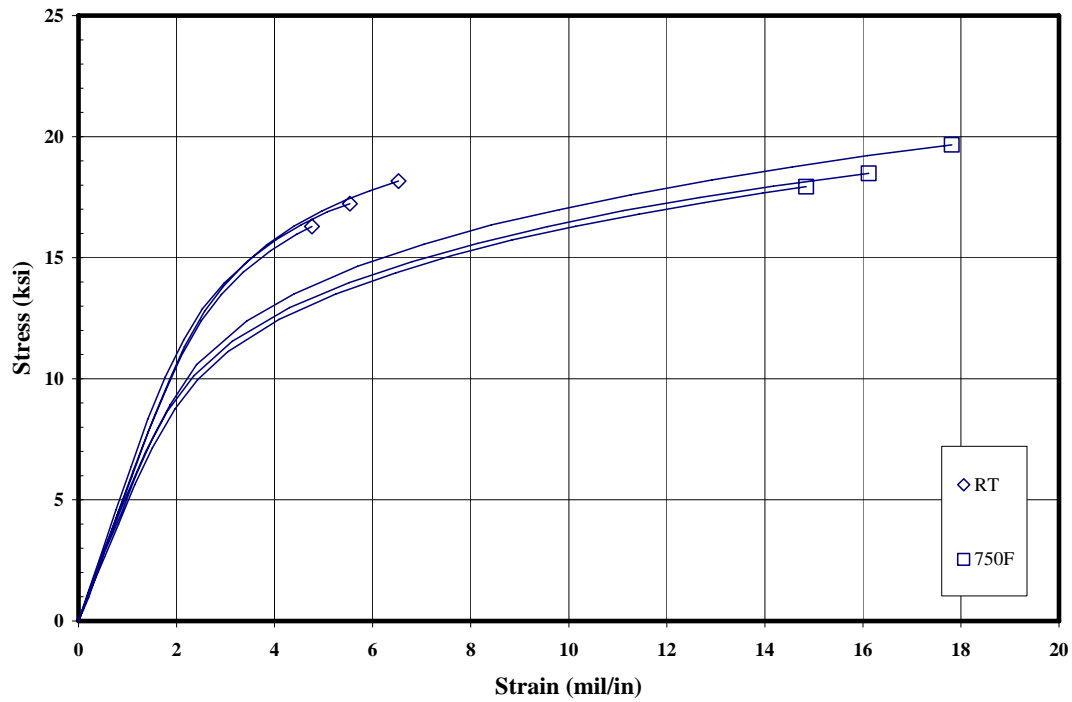


Figure 3.2-15 Axial Stress-Strain Curves for Pall FEAL Filter Element 27065

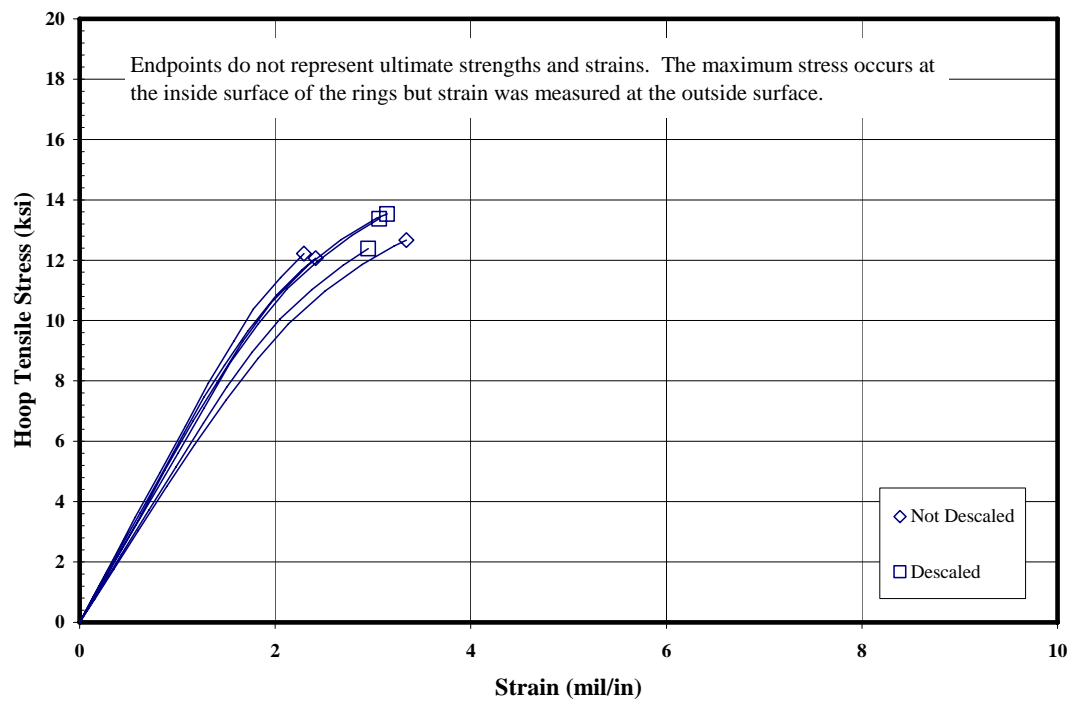


Figure 3.2-16 Hoop Stress-Strain Curves for Pall FEAL Filter Element 27065

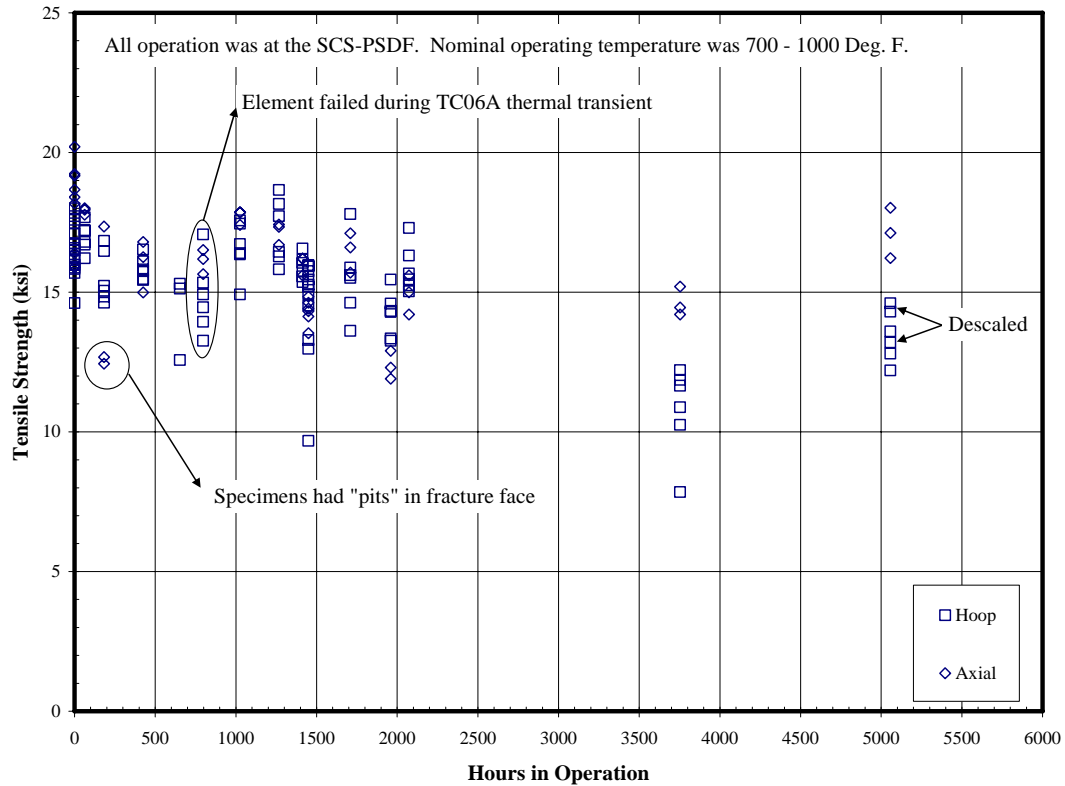


Figure 3.2-17 Ultimate Tensile Strength at RT versus Hours in Operation

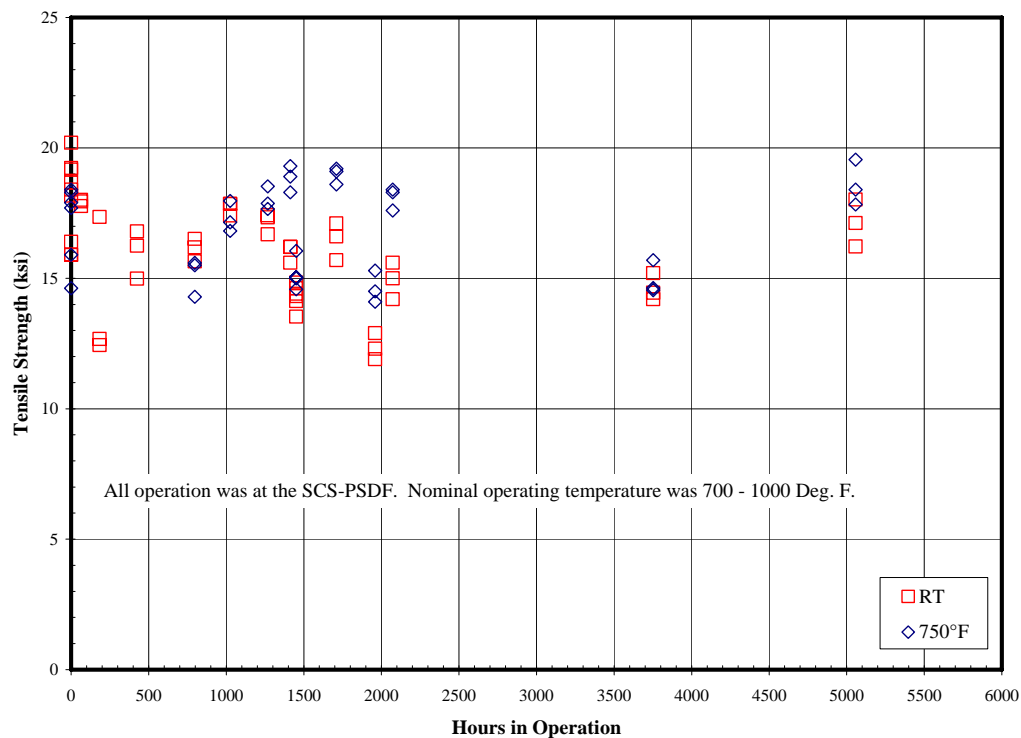


Figure 3.2-18 Ultimate Tensile Strength at RT and 750°F versus Hours in Operation

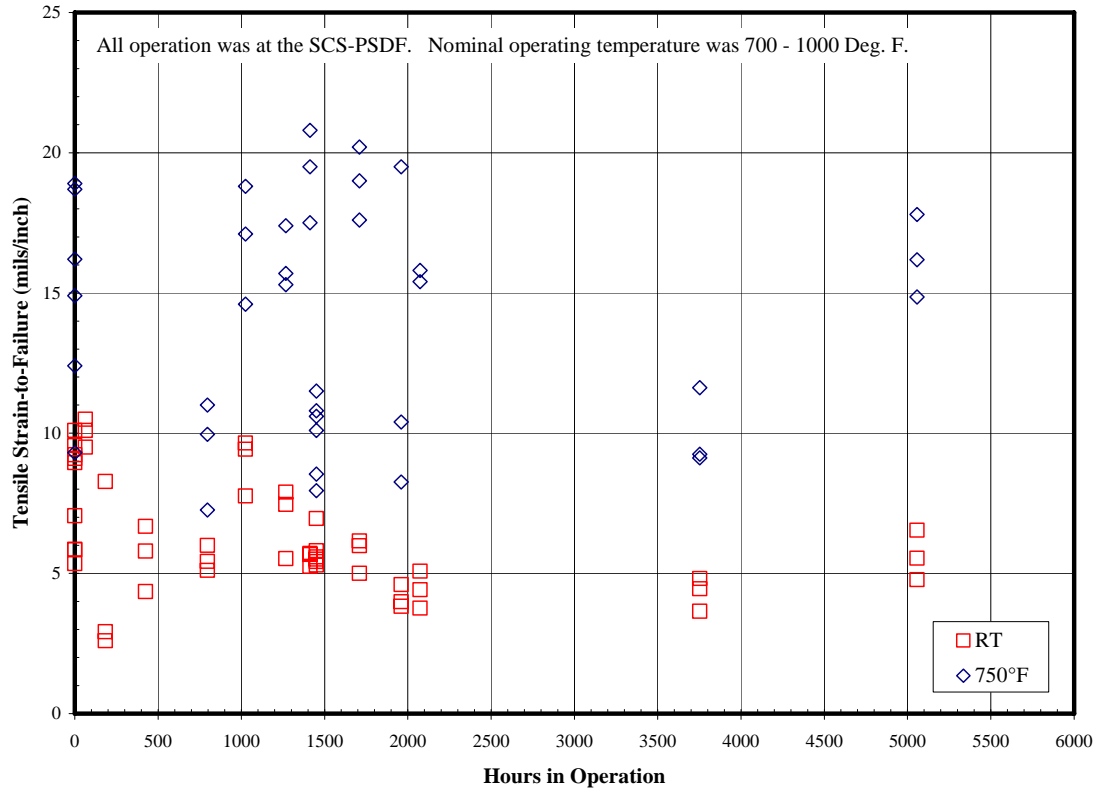


Figure 3.2-19 Strain-to-Failure at RT and 750°F versus Hours in Operation

3.3 PERFORMANCE OF OTHER SYSTEMS

3.3.1 Piloted Syngas Burner/Combustion Turbine

The piloted syngas burner (PSB) operated on syngas for around 7 hours on July 30, 2004, and August 2, 2004, at syngas flow rates up to 13,000 pph. The unit produced 10.3 MWh while on syngas. However, a high frequency noise in the turbine halted subsequent turbine testing temporarily for both simple cycle and syngas operation. Large syngas burners in the past have experienced damage from pressure pulsations initiated by flow induced harmonics or an unstable flame. The SCS Engineering Testing group and Siemens Power determined that destructive pressure pulsations were not present in the PSB combustor, but test conditions did not permit further turbine testing in TC16.

3.3.2 Fuel Cell

A Delphi solid oxide fuel cell operated on syngas generated in the Transport Gasifier for 118 hours throughout TC16. A maximum output of 800 Watts was achieved on July 19. The first fuel cell stack ran for 28 hours, but the performance declined significantly with time. A high concentration of aromatics in the syngas likely contributed to the fast degradation. A new hydrocarbon cracker in the hot gas cleanup train was used for the second fuel cell stack test. The new hydrocarbon cracker slowed the degradation by decreasing the hydrocarbon level from 1000 to 150 ppm. The fuel cell performance only degraded slightly during the first eight hours of testing and then the fuel cell performance remained steady for 82 hours. Although the second test could have continued for a longer period of time, the test stand control system malfunctioned and shut the system down prematurely. This test was a positive step in demonstrating the feasibility of using Solid Oxide Fuel Cell Technology with coal based reformat. For more detailed information on fuel cell testing, refer to the Delphi "Stack Test with Coal Based Syngas Report," DOE-SECA Program DE-FC26-02NT41246.

3.3.3 Steam/Oxygen Eductor

The new steam/oxygen eductor operated very well, blending the steam and oxygen and generating a higher oxygen/steam supply pressure. The gasifier exit pressure ranged between 145 and 155 psig using the standard oxygen/steam mixer. The eductor operated for approximately 330 hours, boosting the oxygen pressure to support operations at gasifier exit pressures from 155 psig to 200 psig (around 215 psig in the lower mixing zone). The eductor operated for approximately 330 hours

3.3.4 Coal Feed Systems

The original coal feeder (FD0210) operated well in TC16, feeding about 1362 tons of coal to the gasifier over a 27 day period. Occasionally, coal fines accumulating in the silo caused the coal feeder to have difficulty transporting the coal through the lock vessel. However, the feeder did not experience problems with the moisture in either the PRB or the lignite coals.

The developmental coal feeder was operated several times during the run. Due to difficulty resulting from plugging in the discharge line, the feeder was transitioned to off-line mode for further testing.

3.3.5 Advanced Syngas Cleanup

The gas cleanup commissioning test run was conducted from July 16, 2004 through August 24, 2004. During the testing, the Transport Gasifier was operating in oxygen blown mode using PRB coal. The gas cleanup unit was commissioned in two modes (bulk cleanup and polishing mode) using syngas for a period of 280 hours. In the bulk cleanup mode, most of the sulfur and HCl were removed in the hot vessels using Syntex sulfur sorbents and Solvay Mineral HCl sorbent respectively. The sorbent bed temperature in the hot vessels mostly ranged from 550 to 750°F at a pressure of 150 psig. In the polishing mode, the syngas exiting the hot vessels was scrubbed in the cold vessels with chemical solutions and water to remove sulfur compounds, trace metals and organic compounds. The cold vessels were operated at ambient temperature and 15 psig pressure. The mini reactor was commissioned in two campaigns utilizing Sud-Chemie catalyst (G-117RR) for ammonia cracking and hydrocarbon reforming over a period of 500 hours. In the first campaign, a mini reactor (1.5 inch ID) was demonstrated in the slip stream of the gas cleanup unit from July 18, 2004 through July 20, 2004. The mini reactor was operated at 1650°F temperature and at 2 to 10 psig pressure. In the second campaign, a new mini reactor (1.687 inch ID) was installed between two hot vessels (RX700B and RX700C). The new mini reactor was demonstrated from July 24, 2004 through August 24, 2004. It was operated at 1600°F temperature and at 135 psig pressure. The gas cleanup unit delivered dry and clean syngas to demonstrate 118 hours of fuel cell testing. The nominal sorbent properties for fixed bed desulfurization, HCl removal, and ammonia cracking are shown in the Tables 3.3-1, 3.3-2, and 3.3-3 respectively. The actual operating parameters for fixed bed desulfurization and HCl removal in hot vessels using syngas, ammonia cracking in mini reactor using bottle gas, and ammonia cracking and hydrocarbon reforming in mini reactor using syngas are shown in Tables 3.3-4, 3.3-5 and 3.3-6 respectively.

The following modifications were completed at the gas cleanup unit during the outage preceding TC16.

- A new hot gas line was installed for fuel cell demonstration test.
- Two new flow meters were installed in the hot and cold gas lines to fuel cell.
- The reactor outlet hot gas sample line was relocated from low pressure side to high pressure side.
- The heat tracing and insulation were added in the downstream of cold vessel, RX704.
- The check valves were moved from the vertical to the horizontal process gas lines.
- The root valves were installed in all utility headers.
- The dirty water tank (TK700) discharge line was modified.
- Three gas sample lines were added to collect gas samples for heavy metal analysis.

During the testing, other modifications were initiated as stated below.

- A two inch mini reactor was installed between the second and the third hot vessels.
- In line filter and vent line were installed in the Fourier transform infrared (FTIR) spectroscopy analyzer gas sample line.

The major accomplishments and observations in TC16 gas cleanup test run included the following:

- The syngas cleanup unit consisting of the hot and cold stream was operated for over 570 continuous hours using nitrogen and syngas combined. During the test, the Transport Gasifier was operating with PRB coal in oxygen blown mode.
- The syngas cleanup test was a 280 hour test run for fixed bed desulfurization and HCl removal in the hot vessels, ammonia cracking and hydrocarbon reforming in the mini reactor, and scrubbing sulfur compounds, trace metals and organic compounds in the cold vessels.
- During the testing, a Delphi solid oxide fuel cell was commissioned using clean syngas for 118 hours. During the first 28 hour fuel cell test, the fuel cell performance declined significantly due to a high concentration of hydrocarbons in the syngas. A second test of 90 hours was performed with a new fuel cell stack. Before the second test, a new mini reactor was installed between the second and the third hot vessels. A nickel-based catalyst was used in the mini reactor to reduce the hydrocarbons in the syngas.
- The hot stream was used for bulk cleaning mode. In the hot stream, the first hot vessel (RX700A) utilized a Syntex catalyst (Puraspec 2010) as a high temperature sulfur sorbent, the second hot vessel (RX700B) utilized a Syntex catalyst (Puraspec 2010) as a high temperature sulfur sorbent and Solvay Mineral catalyst (T-50) as a HCL sorbent, and the third hot vessel (RX700C) utilized a Syntex catalyst (Puraspec 2020) as a low temperature sulfur sorbent. The sorbent bed temperature ranged from 550 to 750°F at a pressure of 150 psig. The H₂S concentration at the inlet of the hot vessels was 300 ppm and the outlet was below the detectible limit of GC (less than 1 ppm). The Syntex catalyst favored the water gas shift reaction in the hot vessels.
- The cold stream was used for polishing mode. The syngas was scrubbed with water in the gas scrubber (RX701). The scrubber water was clean and very little organic crystal deposition was observed in the knock-out pot (RX701B). Cold vessels utilized chemical solutions and water. Sodium hypo-chlorite solution was used in the first cold vessel (RX7002) to remove the trace amount of HCl. Zinc acetate solution was used in the cold vessel, RX7003 to remove the trace amount of sulfur compounds. Water was used in the last cold vessel (RX7004) for a final wash. The cold vessels were operated at ambient temperature and 15 psig pressure. The exit sulfur concentration level was below the ppb range and the HCl concentration level was below 1 ppm.
- The mini reactor was commissioned in two campaigns utilizing Sud-Chemie catalyst (G-117RR) for ammonia cracking and hydrocarbon reforming over a period of 500 hours. In the first campaign, a mini reactor (1.5 inch ID) was demonstrated in the slip stream of the gas cleanup unit. The mini reactor was operated at 1650°F temperature and at 2 to 10 psig. The condensate samples collected in the inlet and outlet of the mini reactor were clear.
- In the second campaign, a new mini reactor (1.687 inch ID) was installed between two hot vessels (RX700B and RX700C). Sud-Chemie nickel-based catalyst, G-117RR was used in the mini reactor to reduce the hydrocarbon concentration in the syngas for fuel cell testing. It was operated at 1600°F temperature and at 135 psig pressure. The gas cleanup unit delivered dry and clean syngas to demonstrate 118 hours of fuel cell testing. It was observed that hydrocarbon such as benzene in the syngas was significantly reduced from 1000 ppm to below 200 ppm. The mini reactor was operating at a higher space velocity due to limited reactor size and higher syngas flow rate for fuel cell testing. After completion of fuel cell

testing, the mini reactor was continued to run along with the hot vessels at lower space velocity. FTIR was used to monitor ammonia concentration. At 200 psig pressure, inlet ammonia measured in the range between 1500 to 2000 ppm, while the outlet ammonia was 200 to 800 ppm. The Sud-Chemie nickel-based catalyst, G-117RR favored the water-gas shift reaction in the mini reactor.

- The electrical band heaters around the hot vessels, RX700A/B/C worked well to control the heating and cooling during testing. A steady vessel temperature up to 1000°F was maintained during the test run.
- The preheater (HX-3) did not perform as expected. The syngas temperature inlet to the hot vessels was limited to 600°F primarily due to high heat loss in the outlet piping at lower syngas flow.
- The gas sampling system in the high pressure side performed well. Liquid back flow from the cold unit to the gas sample line was eliminated. However, intermittent condensation problems were still observed in the GC and FTIR sample lines due to a high moisture content in the syngas during oxygen blown operation.
- All cold stream vessels operated smoothly. The heat tracing in the cold stream was maintained at around 300°F to eliminate condensation in the process lines.

Table 3.3-1 Fixed Bed Desulfurization Sorbents in Hot Vessels

Syntex Sorbent	Zinc Oxide wt %	Physical Properties		
		Shape	Size mm	Density lb/ft ³
Puraspec 2010 (high temp)	84-91	Spheres	5	53
Puraspec 2020 (low temp)	84-91	Spheres	5	72

Table 3.3-2 Fixed Bed HCl Sorbent in Hot Vessel

Chemical Composition	wt %
Na ₂ CO ₃ . NaHCO ₃ . 2H ₂ O	98
SiO ₂	<0.4
H ₂ O Insoluble	2
Physical Properties	
Shape	sphere
Density, lb/ft ³	53
Size Distribution	
U.S. Mesh Screen Size	Cumulative wt %
+20	34
+40	68
+100	94
+140	98
-140	2

Table 3.3-3 Fixed Bed Ammonia Cracking and Hydrocarbon Reforming Catalyst in Mini Reactor

Chemical Composition	wt %
Magnesium Oxide	75 - 90
Nickel Oxide	5 - 15
Calcium Oxide	1 - 5
Aluminum Oxide	1 - 5
Physical Properties	
Shape	Rings
Size (mm)	3 - 4
Density (lb/ft ³)	55 - 75

Table 3.3-4 Actual Operating Parameters for Fixed Bed Desulfurization / HCl removal in Hot Vessels using Syngas

Reactor	Reactor Size ID X Height	Reactor Material	Sorbent Vendor	Sorbent	Sorbent Bed Mass lb	Sorbent Bed Height in	Temperature °F
RX700A	5.187"X 5'	310 SS	Synetix	Puraspec 2010	37	42	750
RX700B	5.187"X 5'	310 SS	Syntex/Solvey Mineral	Puraspec 2010/T- 50	30 / 5	40	750
RX700C	5.187"X 4'	CS	Synetix	Puraspec 2020	30	36	550

Table 3.3-5 Actual Operating Parameters for Fixed Bed Ammonia Cracking in Mini Reactor Using Bottle Gases

Test Campaign I	
Duration	07/18/2004 to 07/20/2004
Mini Reactor	RX-301 (old)
Mini Reactor Size	1.5" ID x 4'Ht
Sorbent	Sud-Chemie G-117RR
Sorbent bed mass, lb	0.3
Sorbent bed height, in	5
Syngas flow rate, lb/hr	1
Pressure, psig	2 - 10
Temperature, °F	1650
NH3 conversion, %	>95%

Table 3.3-6 Actual Operating Parameters for Fixed Bed Ammonia Cracking and Hydrocarbon Reforming in Mini Reactor using Syngas

Test Campaign II	
Duration	07/24/2004 to 08/24/2004
Mini Reactor	RX-301 (new)
Mini Reactor Size	1.687" ID x 4' Ht
Sorbent	Sud-Chemie G-117RR
Sorbent bed mass, lb	3
Sorbent bed height, in	40
Syngas flow rate, lb/hr	10 - 20
Pressure, psig	135 - 200
Temperature, °F	1600
Benzene Conversion, %	80 - 95
NH3 conversion, %	60 - 85

APPENDIX A1 OPERATION HISTORY

Conversion of the Transport Reactor train to gasification mode of operation was performed from May to September 1999. The first gasification test run, GCT1, was a 233-hour test run to commission the Transport Gasifier and to characterize the limits of operational parameter variations. GCT1 was started on September 9, 1999, with the first part completed on September 15, 1999. The second part of GCT1 was started on December 7, 1999, and completed on December 15, 1999. This test run provided the data necessary for preliminary analysis of gasifier operations and for identification of necessary modifications to improve equipment and process performance. Five different feed combinations of coal and sorbent were tested to gain a better understanding of the gasifier solids collection system efficiency.

GCT2, a 218-hour characterization test run, was started on April 10, 2000, and completed on April 27, 2000. Additional data was taken to analyze the effect of different operating conditions on gasifier performance and operability. A blend of several Powder River Basin (PRB) coals was used with Longview limestone from Alabama. In the outage following GCT2, the Transport Gasifier was modified to improve the operation and performance of the gasifier solids collection system. The most fundamental change was the addition of the loop seal underneath the primary cyclone.

GCT3 was a 184-hour characterization with the primary objective to commission the loop seal. A hot solids circulation test (GCT3A) was started on December 1, 2000, and completed December 15, 2000. After a one-month outage to address maintenance issues with the main air compressor, GCT3 was continued. The second part of GCT3 (GCT3B) was started on January 20, 2001, and completed on February 1, 2001. During GCT3B, a blend of several PRB coals was used with Bucyrus limestone from Ohio. The loop seal performed well allowing much higher solids circulation rates and higher syngas heating values. Also, the improved collection efficiency of the cyclone resulted in lower relative solids loading to the PCD and higher carbon conversion.

GCT4, a 242-hour characterization test run, was started on March 7, 2001, and completed on March 30, 2001. A blend of several PRB coals with Bucyrus limestone from Ohio was used. More experience was gained with the loop seal operations and additional data was collected to better understand gasifier performance.

TC06, a 1025-hour test campaign, was started on July 4, 2001, and completed on September 24, 2001. A blend of several PRB coals with Bucyrus limestone from Ohio was used. Due to its length and stability of operation, the TC06 test run provided valuable data necessary to analyze long term gasifier operations and to identify necessary modifications to improve equipment and process performance, as well as progressing the goal of many thousands of hours of filter element exposure.

TC07, a 442-hour test campaign, was started on December 11, 2001, and completed on April 5, 2002. A blend of several PRB coals and a bituminous coal from the Calumet mine in Alabama were tested with Bucyrus limestone from Ohio. Due to operational difficulties with the gasifier (stemming from instrumentation problems) the unit was taken offline several times. PCD operations were relatively stable considering the numerous gasifier upsets.

TC08 was a 365-hour test campaign to commission the gasifier in oxygen-blown mode of operation. TC08 started on June 9, 2002, and completed on June 29, 2002. A blend of several PRB coals was

tested in air-blown, enriched air- and oxygen-blown modes of operation. The transition from different modes of operation was smooth and it was demonstrated that the transition from air to oxygen could be made within 15 minutes. Both gasifier and PCD operations were stable during the test run.

TC09 was a 309-hour test campaign to characterize the gasifier and PCD operations in air- and oxygen-blown mode of operations using a bituminous coal from the Sufco mine in Utah. TC09 was started on September 3, 2002, and completed on September 26, 2002. Both gasifier and PCD operations were stable during the test run.

TC10 was a 416-hour test campaign to conduct long-term tests to evaluate the gasifier and PCD operations in oxygen-blown mode of operations using a blend of several PRB coals. TC10 was started on November 16, 2002, and completed on December 18, 2002. Despite problems with the coal mills, coal feeder, pressure tap nozzles and the standpipe, the gasifier did experience short periods of stability during oxygen blown operations. During these periods, the syngas quality was high. During TC10, over 609 tons of PRB subbituminous coals were gasified.

TC11 was a 192-hour test campaign to conduct short-term tests to evaluate the gasifier and PCD operations in air and oxygen-blown mode of operations using Falkirk lignite from North Dakota. TC11 was started on April 7, 2003, and completed on April 18, 2003. During TC11, the lignite proved difficult to feed due to difficulties in the mill operation as a result of the high moisture content in the fuel. However, the gasifier operated well using lignite, with high circulation rates, riser densities and even temperature profiles. Consequently, the temperature distribution in both the mixing zone and the riser was more uniform than in any previous test run, varying less than 10°F throughout the gasifier.

TC12 was a 733-hour test campaign to conduct short-term tests to evaluate the gasifier and PCD operations in air- and oxygen-blown mode of operations using a blend of several PRB coals. TC12 was started on May 16, 2003, and completed on July 14, 2003. A primary focus for TC12 was the commissioning of a new gas cleanup system and operating a fuel cell on syngas derived from the Transport Gasifier. The fuel cell system and gas cleanup system both performed well during the testing.

TC13 was a 501-hour test campaign to conduct short-term tests to evaluate gasifier, PSB, and PCD operations in air-blown mode of operations using a blend of several PRB coals, as well as to conduct short-term tests to evaluate gasifier and PCD operations using two different types of lignite from the Freedom Mine in North Dakota. One type of lignite had high ash sodium content, while the other types had low ash sodium content. TC13 was started on September 30, 2003, and completed on November 2, 2003. The syngas-to-PSB testing lasted for a total of about 6 hours. While successful, the hydraulic system on the turbine cranking motor failed and prevented further PSB testing. The low sodium lignite testing went well, but lowering the gasifier temperature to below 1500°F was necessary to prevent ash agglomeration with the high sodium lignite.

TC14 was a 214-hour test campaign to conduct short-term tests to evaluate the gasifier, PSB, and PCD operations in air and oxygen blown mode of operations using a blend of several PRB coals. TC14 began on February 16, 2004, and ended on February 28, 2004. The syngas-to-PSB testing lasted for a total of about 17 hours at syngas flow rates up to 17,000 pph, contributing about 82 percent of the total energy to the PSB. The Continuous Fine Ash Depressurization (CFAD) unit was commissioned during TC14. The new system worked well and operated for 190 hours. The gasifier operation was smooth, with the exception of the decrease in the primary cyclone efficiency which caused the gasifier to continuously lose bed material.

TC15 was a 200-hour test campaign to conduct short-term tests to evaluate the gasifier, PSB and PCD operations in air and oxygen blown modes of operations using a blend of several PRB coals. TC15 began on April 19, 2004, and ended on April 29, 2004. The syngas-to-PSB testing lasted for approximately 15 hours at syngas flow rates up to 17,000 pph, contributing about 86 percent of the total energy to the PSB. The gasifier experienced stable operations in air-blown mode and less stable operations in oxygen-blown mode due to poor solids circulation. A primary focus of TC15 was to commission and test modifications made to the syngas cleanup system. The system was effective in reducing the sulfur content and achieving an absorption capacity between 19 and 30 percent.

TC16, the subject of this report, was an 835-hour test campaign to conduct short-term tests to evaluate gasifier, PSB, and PCD operations in air- and oxygen-blown modes of operations using a blend of several PRB coals, as well as to conduct short-term tests to evaluate gasifier and PCD operations using high sodium lignite coal from the Freedom Mine in North Dakota. TC16 began on July 14, 2004, and ended on August 24, 2004. The syngas-to-PSB testing lasted for approximately seven hours at syngas flow rates up to 13,000 pph and the combustion turbine (CT) operated for about 20 hours. A Delphi solid oxide fuel cell operated on syngas for 118 hours during TC16. The first fuel cell stack ran for 28 hours, during which time the performance declined significantly. Another fuel cell stack was installed and fuel cell performance only degraded slightly during the first eight hours of testing, then remained steady for 82 hours. The new steam/oxygen educator operated very well, generating steam/oxygen supply pressures up to 200 psig (215 psig in the lower mixing zone). Operations with Freedom lignite testing in TC16 was markedly improved over the testing done in TC13.

Figure A1-1 gives a summary of operating test hours achieved with the gasification process at the PSDF.

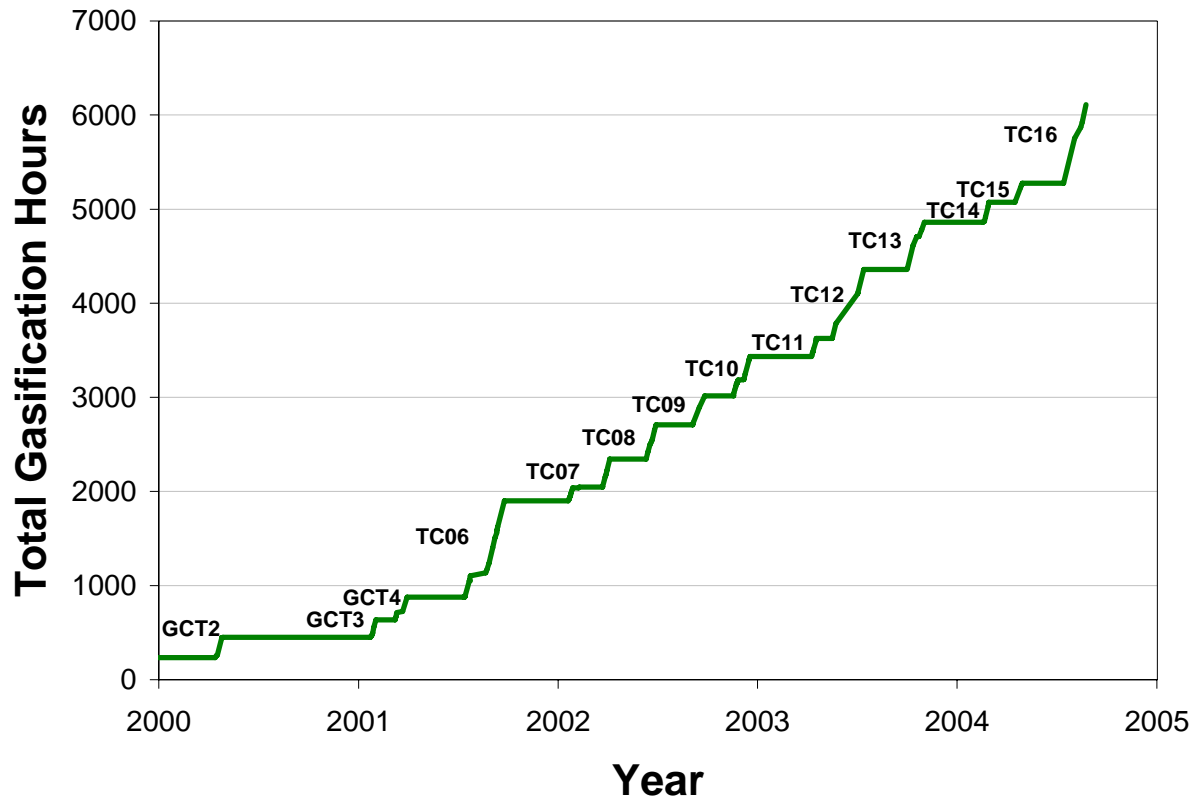


Figure A1-1 Operating Hours Summary for the PSDF Gasification Process

APPENDIX A2 EQUIPMENT LIST

Major Equipment in the Transport Gasifier Train

TAG NAME	DESCRIPTION
	Delphi Solid Oxide Fuel Cell
BR0201	Gasifier Start-Up Burner
BR0401	Atmospheric Syngas Combustor Oxidizer
BR0452	Piloted Syngas Burner
BR0602	Atmospheric Fluidized Bed Combustor (AFBC) Start-Up Burner
CO0201	Main Air Compressor
CO0451	Turbine Air Compressor
CO0601	AFBC Air Compressor
CY0201	Primary Cyclone in the Gasifier Loop
CY0207	Disengager in the Gasifier Loop
CY0601	AFBC Cyclone
DR0402	Steam Drum
DY0201	Feeder System Air Dryer
FD0206	Spent Solids Screw Cooler
FD0210	Coal Feeder System
FD0220	Sorbent Feeder System
FD0502	Fines Screw Cooler
FD0510	Spent Solids Transporter System
FD0520	Fines Transporter System
FD0530	Spent Solids Feeder System
FD0540	Continuous Fine Ash Depressurization (CFAD) System
FD0602	AFBC Solids Screw Cooler
FD0610	AFBC Sorbent Feeder System
FL0301	PCD
FL0401	Compressor Intake Filter
GN0451	Turbine Generator
GT0451	Gas Turbine
HX0202	Primary Gas Cooler
HX0204	Transport Air Cooler
HX0402	Secondary Gas Cooler
HX0540	CFAD Collection Drum/Heat Exchanger
HX0601	AFBC Heat Recovery Exchanger
ME0540	Heat Transfer Fluid System
RX0201	Transport Gasifier
SI0602	Spent Solids Silo
SU0601	AFBC

Major Equipment in the Balance of Plant (page 1 of 3)

TAG NAME	DESCRIPTION
B02920	Auxiliary Boiler
B02921	Auxiliary Boiler – Superheater
CL2100	Cooling Tower
C02201A-D	Service Air Compressor A-D
C02202	Air-Cooled Service Air Compressor
C02203	High-Pressure Air Compressor
C02601A-C	Reciprocating N ₂ Compressor A-C
CR0104	Coal and Sorbent Crusher
CV0100	Crushed Feed Conveyor
CV0101	Crushed Material Conveyor
DP2301	Baghouse Bypass Damper
DP2303	Inlet Damper on Dilution Air Blower
DP2304	Outlet Damper on Dilution Air Blower
DY2201A-D	Service Air Dryer A-D
DY2202	Air-Cooled Service Air Compressor Air Dryer
DY2203	High-Pressure Air Compressor Air Dryer
FD0104	MWK Coal Transport System
FD0105	FW Coal Transport System
FD0111	MWK Coal Mill Feeder
FD0112	FW Coal Mill Feeder
FD0113	Sorbent Mill Feeder
FD0140	Coke Breeze and Bed Material Transport System
FD0154	MWK Limestone Transport System
FD0810	Ash Unloading System
FD0820	Baghouse Ash Transport System
FL0700	Baghouse
FN0700	Dilution Air Blower
H00100	Reclaim Hopper
H00105	Crushed Material Surge Hopper
H00252	Coal Surge Hopper
H00253	Sorbent Surge Hopper
HT2101	MWK Equipment Cooling Water Head Tank
HT2103	SCS Equipment Cooling Water Head Tank
HT0399	60-Ton Bridge Crane
HX2002	MWK Steam Condenser
HX2003	MWK Feed Water Heater

Major Equipment in the Balance of Plant (page 2 of 3)

TAG NAME	DESCRIPTION
HX2004	MWK Subcooler
HX2103A	SCS Cooling Water Heat Exchanger
HX2103B	FW Cooling Water Heat Exchanger
HX2103C	MWK Cooling Water Heat Exchanger
LF0300	Propane Vaporizer
MC3001-3017	MCCs for Various Equipment
ME0700	MWK Stack
ME0814	Dry Ash Unloader for MWK Train
ML0111	MWK Coal Mill
ML0112	FW Coal Mill
ML0113	Sorbent Mill for Both Trains
PG0011	Oxygen Plant
PG2600	Nitrogen Plant
PU2000A-B	MWK Feed Water Pump A-B
PU2100A-B	Raw Water Pump A-B
PU2101A-B	Service Water Pump A-B
PU2102A-B	Cooling Tower Make-Up Pump A-B
PU2103A-D	Circulating Water Pump A-D
PU2107	SCS Cooling Water Make-Up Pump
PU2109A-B	SCS Cooling Water Pump A-B
PU2110A-B	FW Cooling Water Pump A-B
PU2111A-B	MWK Cooling Water Pump A-B
PU2300	Propane Pump
PU2301	Diesel Rolling Stock Pump
PU2302	Diesel Generator Transfer Pump
PU2303	Diesel Tank Sump Pump
PU2400	Fire Protection Jockey Pump
PU2401	Diesel Fire Water Pump #1
PU2402	Diesel Fire Water Pump #2
PU2504A-B	Waste Water Sump Pump A-B
PU2507	Coal and Limestone Storage Sump Pump
PU2700A-B	Demineralizer Forwarding Pump A-B

Major Equipment in the Balance of Plant (page 3 of 3)

TAG NAME	DESCRIPTION
PU2920A-B	Auxiliary Boiler Feed Water Pump A-B
SB3001	125-V DC Station Battery
SB3002	Uninterruptible Power Supply (UPS)
SC0700	Baghouse Screw Conveyor
SG3000-3005	4160-V, 480-V Switchgear Buses
SI0101	MWK Crushed Coal Storage Silo
SI0102	FW Crushed Coal Storage Silo
SI0103	Crushed Sorbent Storage Silo
SI0111	MWK Pulverized Coal Storage Silo
SI0112	FW Pulverized Coal Storage Silo
SI0113	MWK Limestone Silo
SI0114	FW Limestone Silo
SI0810	Ash Silo
ST2601	Nitrogen Storage Tube Bank
TK2000	MWK Condensate Storage Tank
TK2001	FW Condensate Tank
TK2100	Raw Water Storage Tank
TK2300A-D	Propane Storage Tank A-D
TK2301	Diesel Storage Tank
TK2401	Fire Water Tank
XF3000A	230/4.16-kV Main Power Transformer
XF3001B-5B	4160/480-V Station Service Transformer No. 1-5
XF3001G	480/120-V Miscellaneous Transformer
XF3010G	120/208 Distribution Transformer
XF3012G	UPS Isolation Transformer
VS2203	High-Pressure Air Receiver

APPENDIX A3 MATERIAL AND ENERGY BALANCES

Material balances are useful in checking the accuracy and consistency of data as well as determining periods of operation where the data is suitable for model development or commercial plant design. Total material balances for each operating period are given in Figure A3-1 which compare the total mass in and the total mass out. The overall mass balance was good, with all of the relative differences at ± 10 percent. The relative difference (relative error) is defined as the Transport Gasifier feeds minus the products divided by the feeds ($\{\text{In-Out}\}/\text{In}$).

The main contributors to the material balance are the syngas flow rate (13,900 to 24,200 pph), the air flow rate (900 to 13,600 pph), the oxygen flow rate (0 to 2,600 pph), the steam flow rate (500 to 4,000 pph), the nitrogen flow rate (5,400 to 6,500 pph), and the coal feed rate (2,400 to 4,500 pph). Although higher than in previous test runs, the sorbent feed rate (0 to 300 pph) contributed only a small amount to the overall mass balance. Of the total air-blown operating periods using PRB and lignite, all but one of them had higher overall mass flow rates than the oxygen-blown operating periods using PRB due to additional nitrogen from air fed to the Transport Gasifier.

The TC16 Transport Gasifier energy balance is shown in Figure A3-2 with standard conditions chosen to be a pressure of 1.0 atmosphere and a temperature of 80°F. As shown in the figure, the TC16 energy balances were mostly within ± 10 percent error for PRB coal. However, only about one-half of the lignite operating periods fell within the ± 10 percent error, with the remainder falling within ± 15 percent error. The energy entering the gasifier consisted of the coal, air, and steam fed to the Transport Gasifier. The nitrogen, oxygen and sorbent fed to the gasifier were considered to be at standard conditions (80°F) and, hence, had zero enthalpy. The nitrogen and oxygen feeds actually entered the gasifier at a higher temperature than standard conditions, but compared to the other feed enthalpies, this neglected input energy is insignificant. Since the amount of solids removed from the standpipe was negligible, the energy exiting the gasifier consisted of only the syngas and PCD solids. The analysis used the lower heating value of the coal, the PCD solids, and the syngas.

The energy of the syngas was determined at the Transport Gasifier primary cyclone exit. Since the total syngas flow measurement is located downstream of the PCD, 320 pounds of nitrogen per hour that flowed to the PCD inlet and outlet particulate sampling trains was subtracted from the exit flow rate to determine the actual syngas rate from the cyclone. The sensible enthalpy of the syngas was determined by the overall gas heat capacity from the syngas compositions. The syngas and PCD solids energy consists of both latent and sensible heat. The heat loss from the Transport Gasifier was estimated to be 3.5 million Btu/hr.

The TC16 carbon balance is shown in Figure A3-3. The carbon balance was within ± 15 percent, with the oxygen blown PRB testing better than the air blown PRB and lignite testing. The carbon balance gives a measure of how accurate the TC16 carbon conversions are. The most probably sources of error in the carbon balance are the coal-feed rate measurement and the syngas-flow rate measurement.

The TC16 sulfur balance is shown in Figure A3-4. The sulfur balance was within ± 25 percent, with the oxygen blown PRB testing better than the air blown PRB testing. The lignite air-blown sulfur

balances were not very good with several tests having higher sulfur balance errors than 25 percent. The sulfur balance gives a measure of the accuracy of the TC16 sulfur captures. The most probable sources or error in the sulfur balance is the calculation of the syngas sulfur. The syngas sulfur is calculated by a combustion calculation based on a flue gas sulfur dioxide measurement rather than directly measured from the syngas.

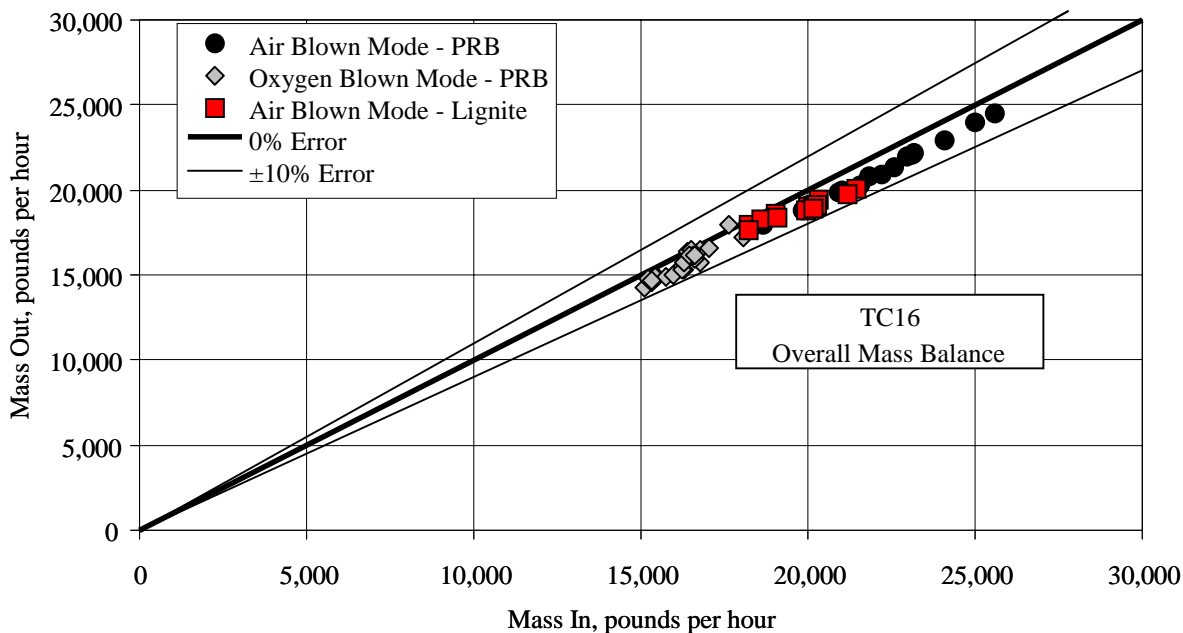


Figure A3-1 Mass Balance

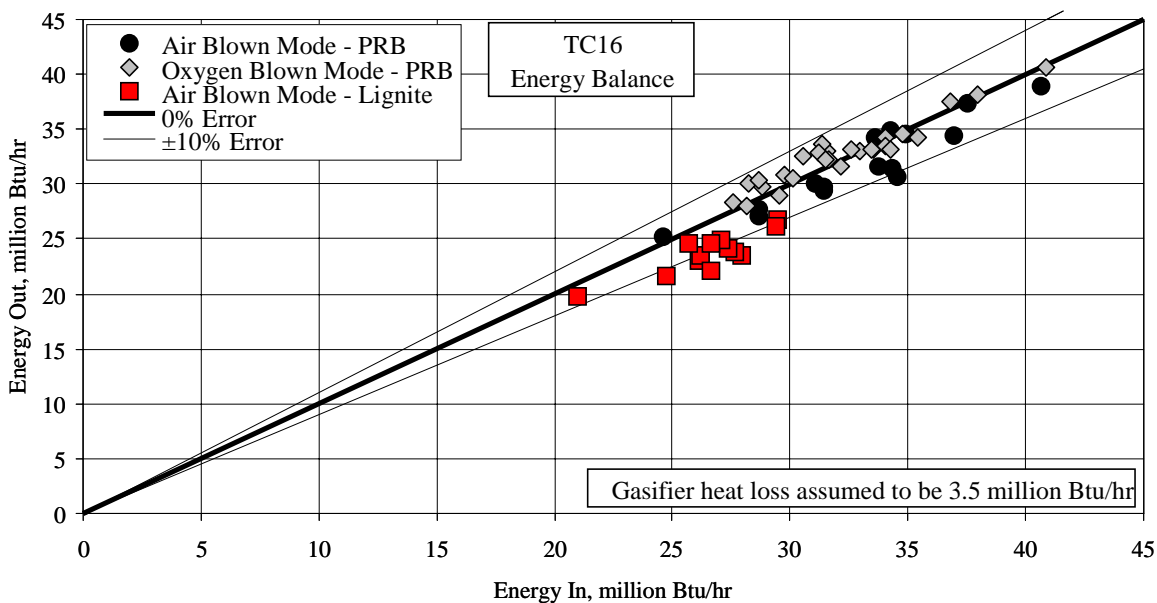


Figure A3-2 Energy Balance

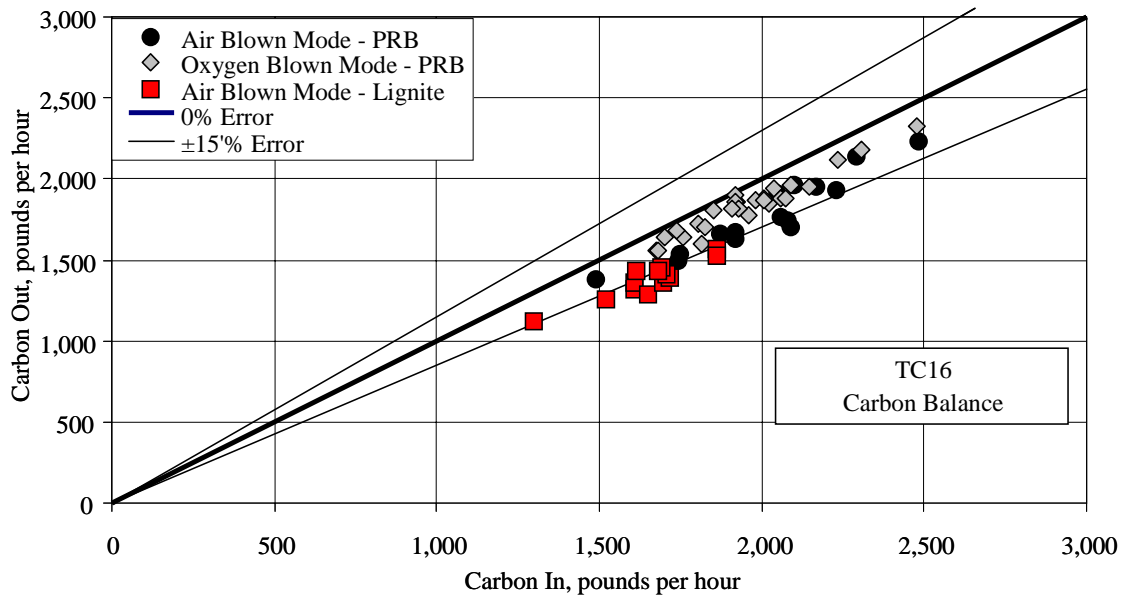


Figure A3-3 Carbon Balance

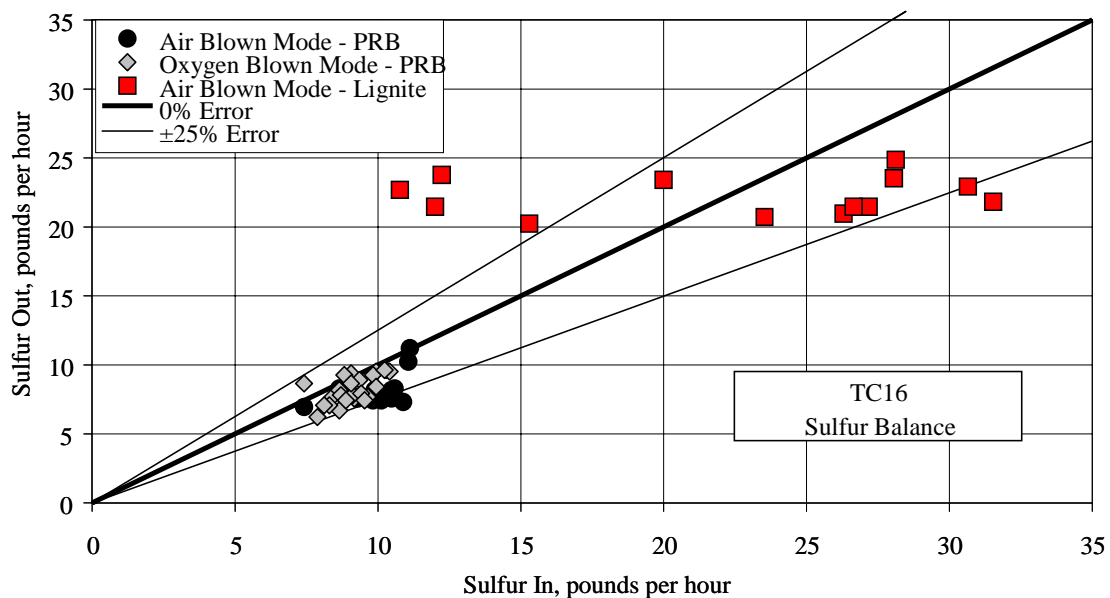


Figure A3-4 Sulfur Balance

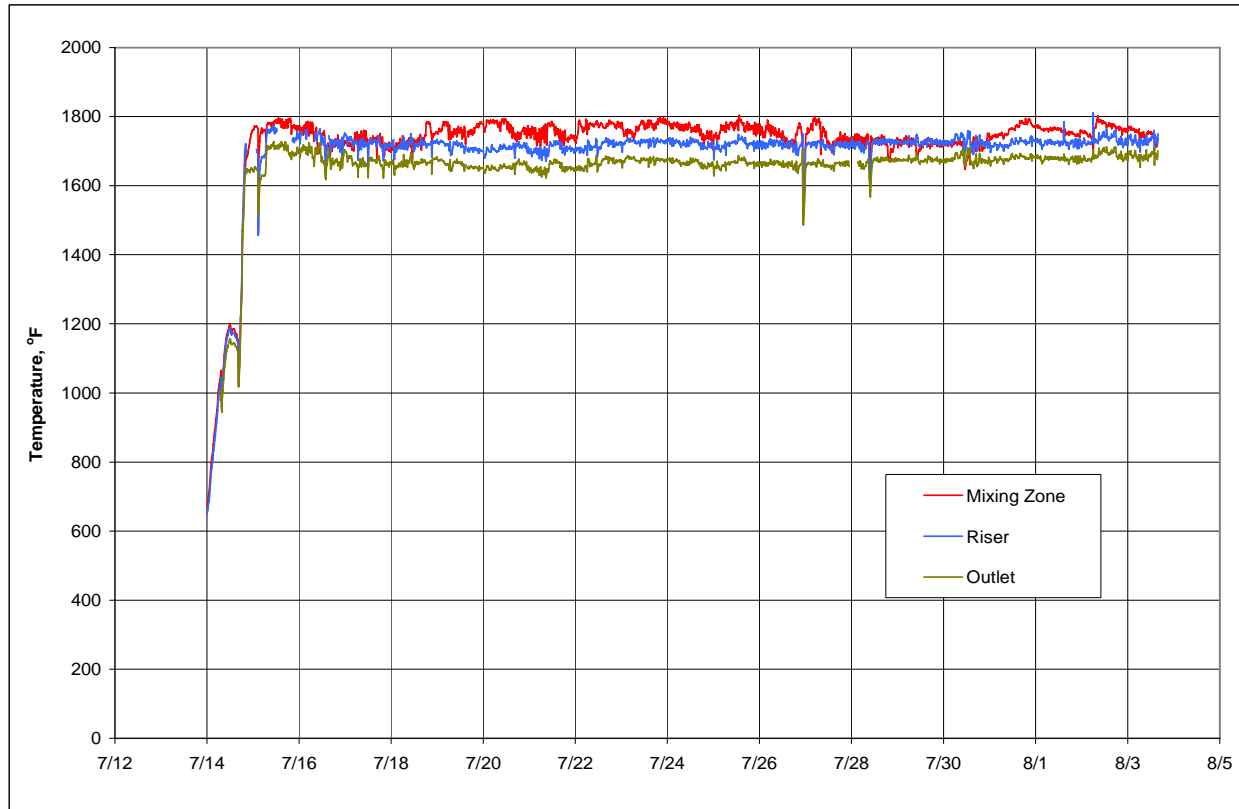


Figure A4-1 Gasifier Mixing Zone, Riser, and Outlet Temperatures, 7/14/04 through 8/4/04

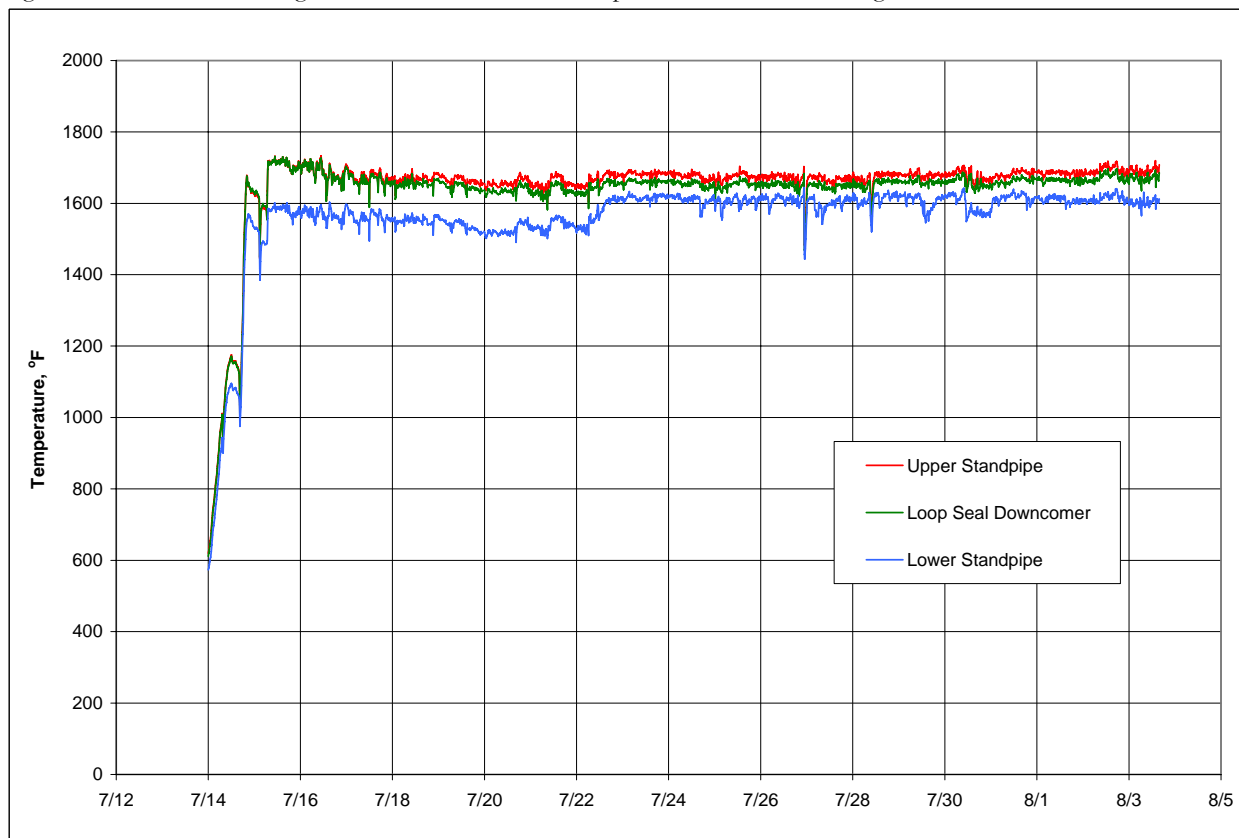


Figure A4-2 Standpipe and Loop Seal Temperatures, 7/14/04 through 8/4/04

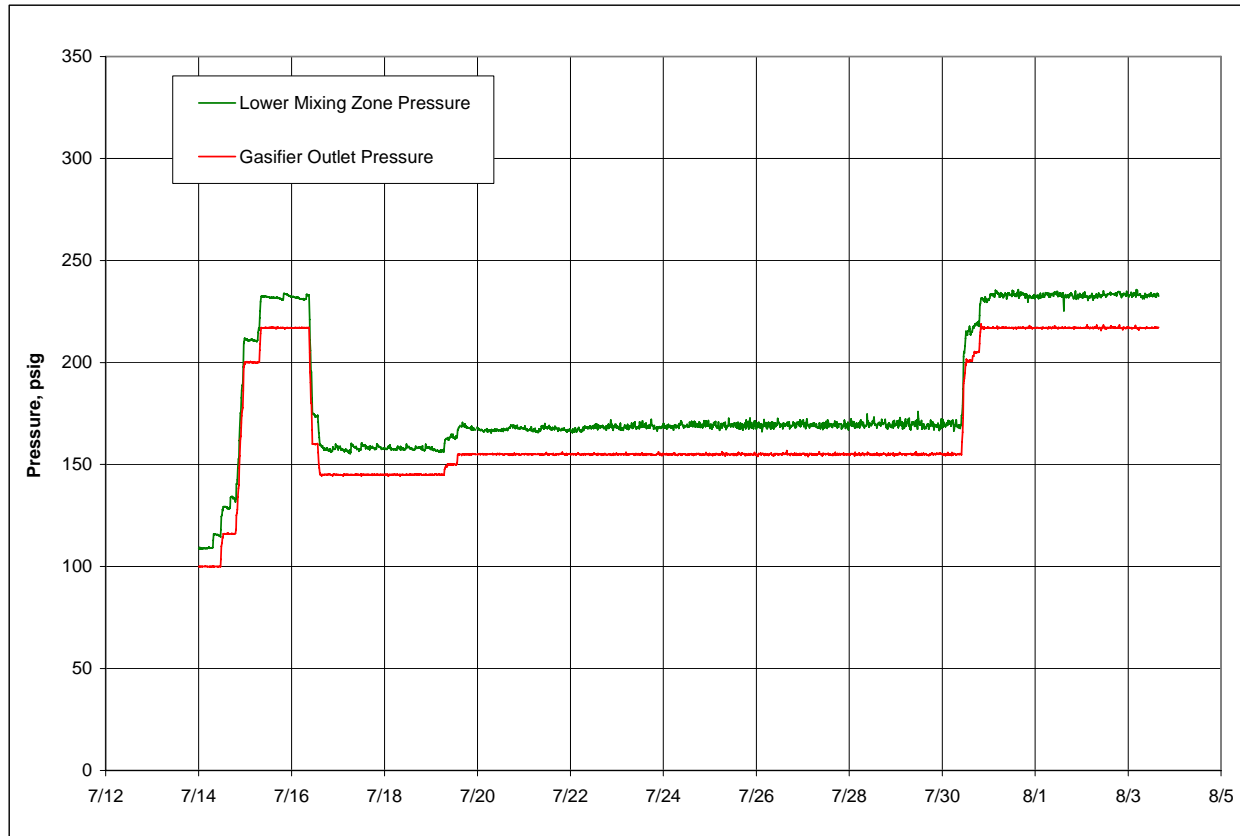


Figure A4-3 Gasifier Pressures, 7/14/04 through 8/4/04

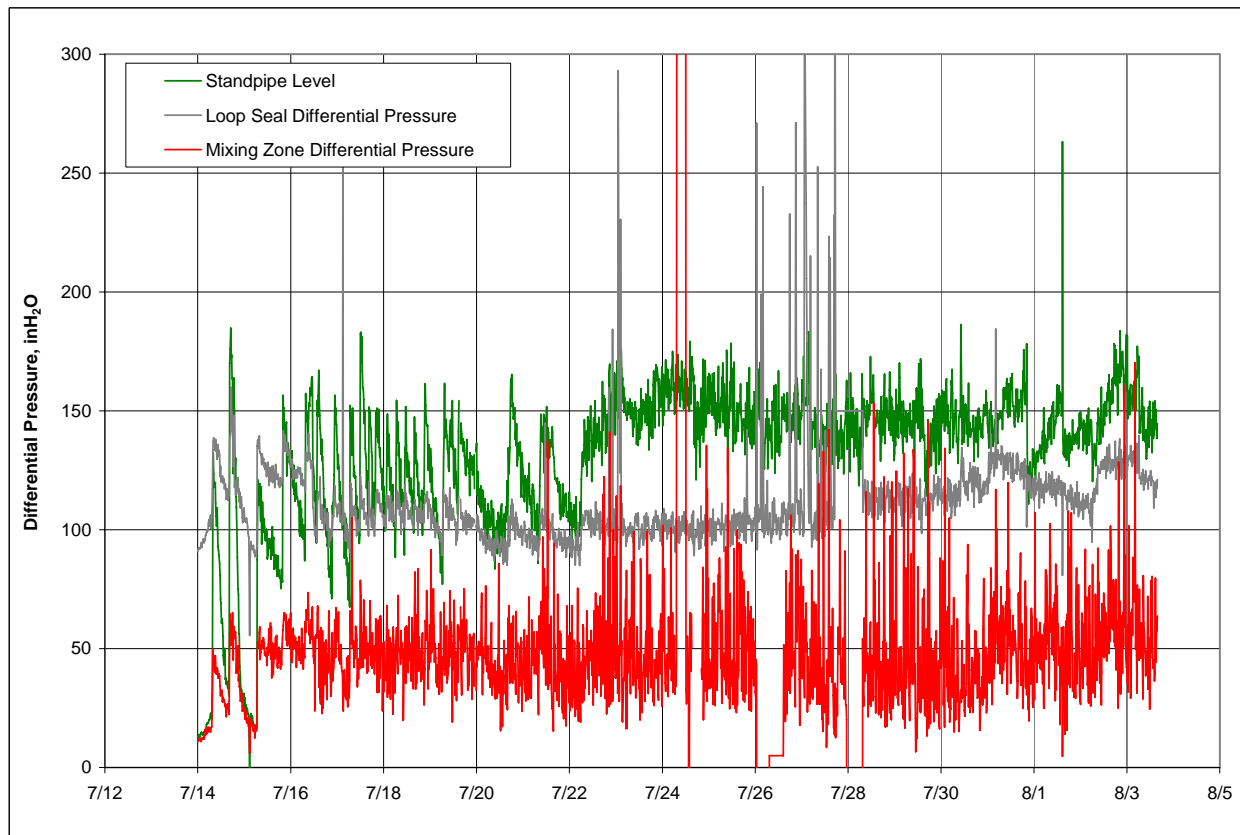


Figure A4-4 Gasifier Differential Pressures, 7/14/04 through 8/4/04

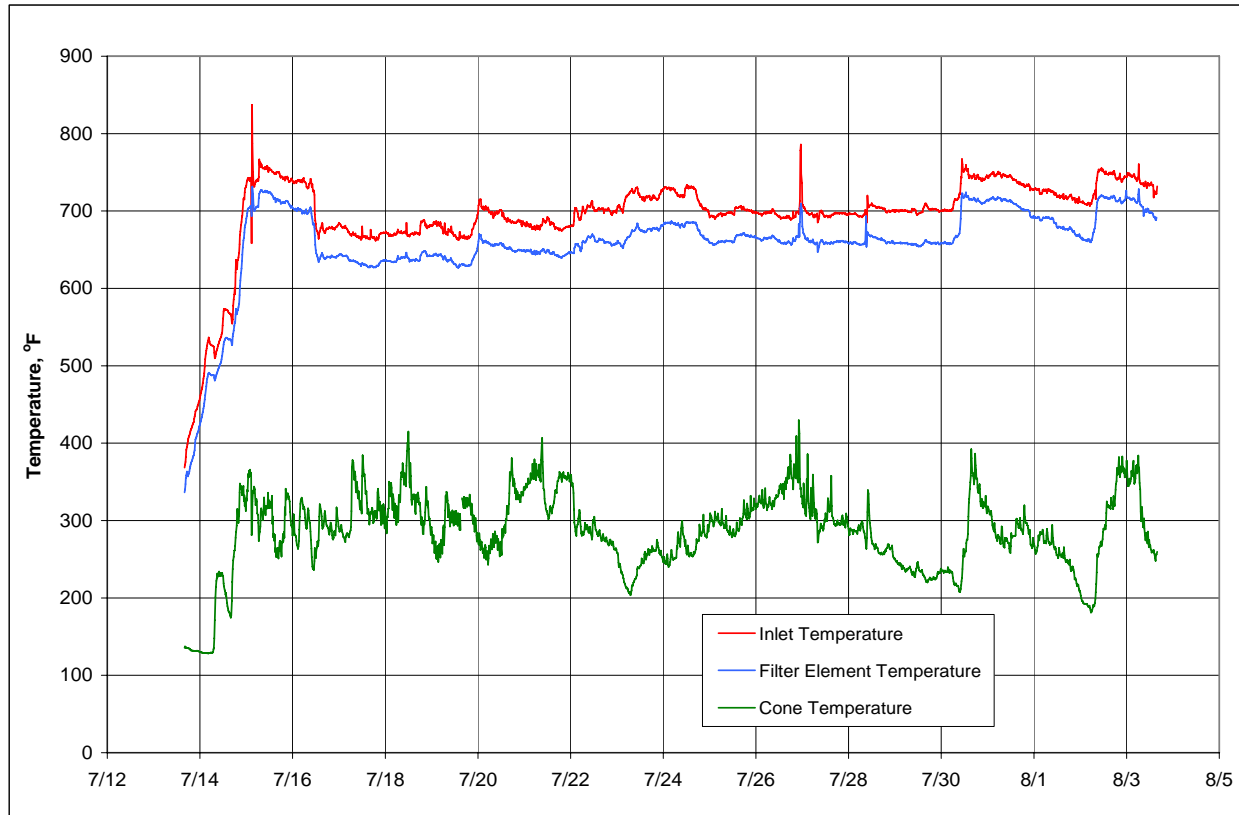


Figure A4-5 PCD Temperatures, 7/14/04 through 8/4/04

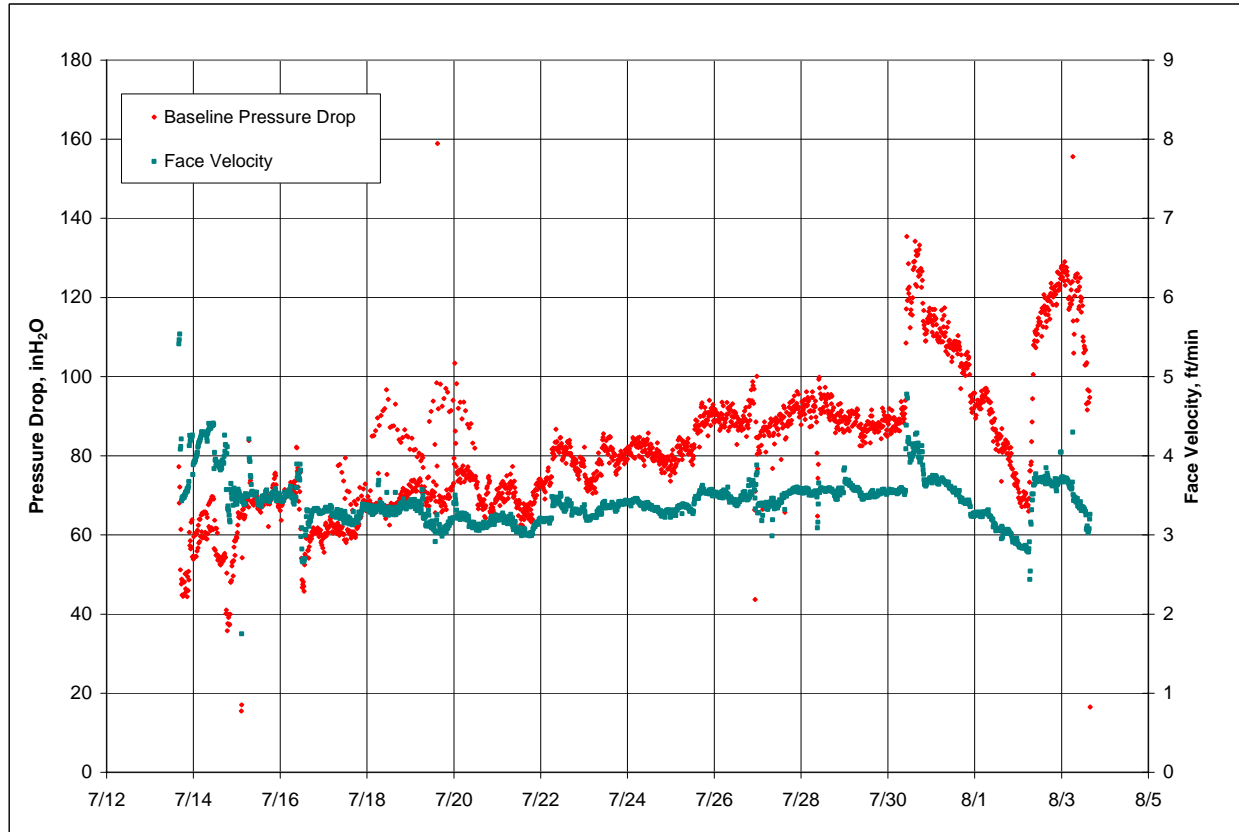


Figure A4-6 PCD Baseline Pressure Drop and Face Velocity, 7/14/04 through 8/4/04

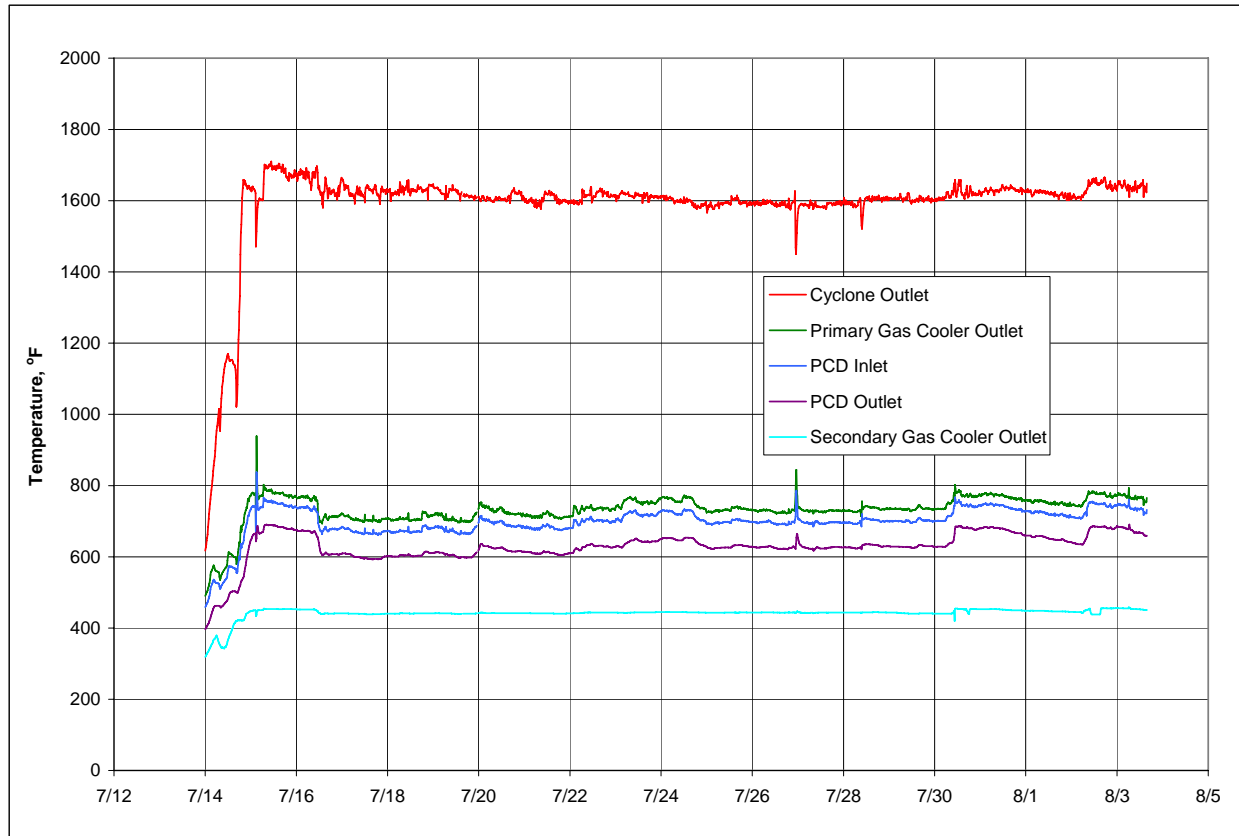


Figure A4-7 System Temperature Profile, 7/14/04 through 8/4/04

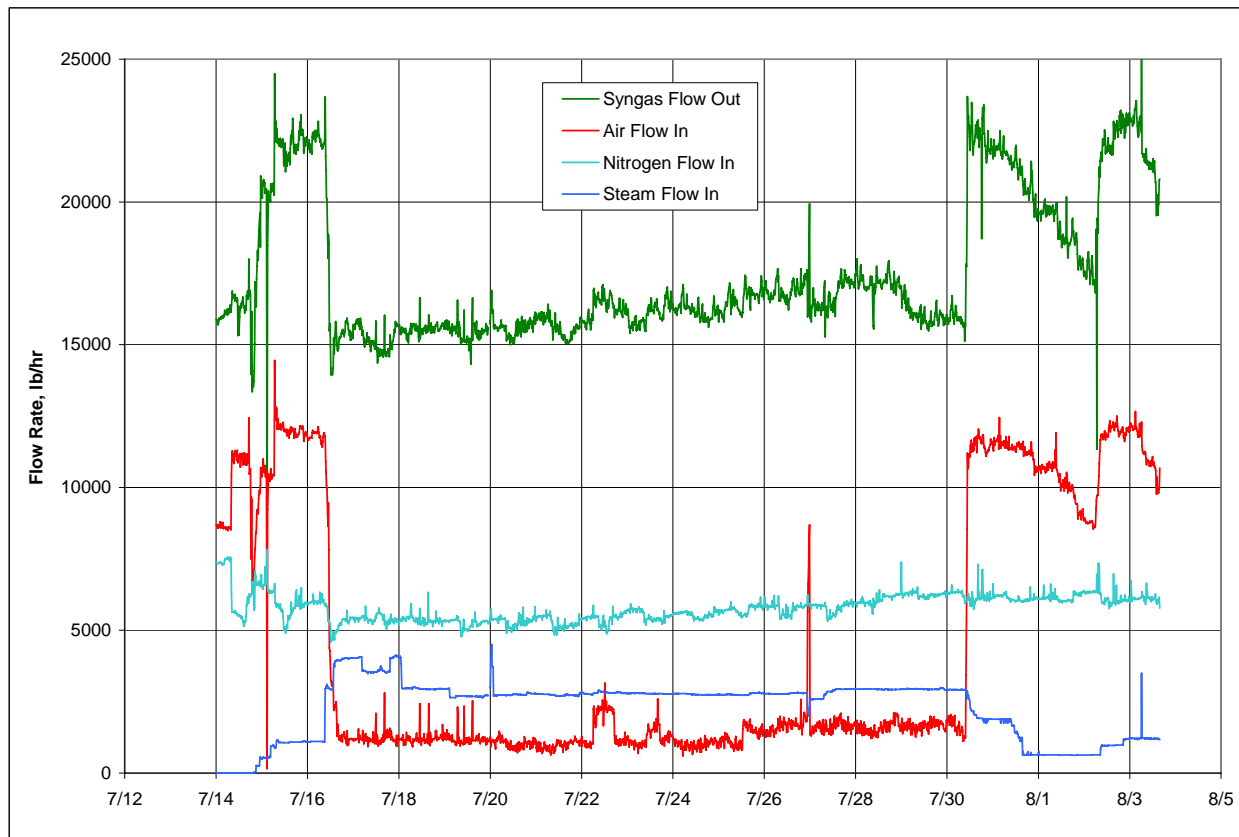


Figure A4-8 System Gas Flows, 7/14/04 through 8/4/04

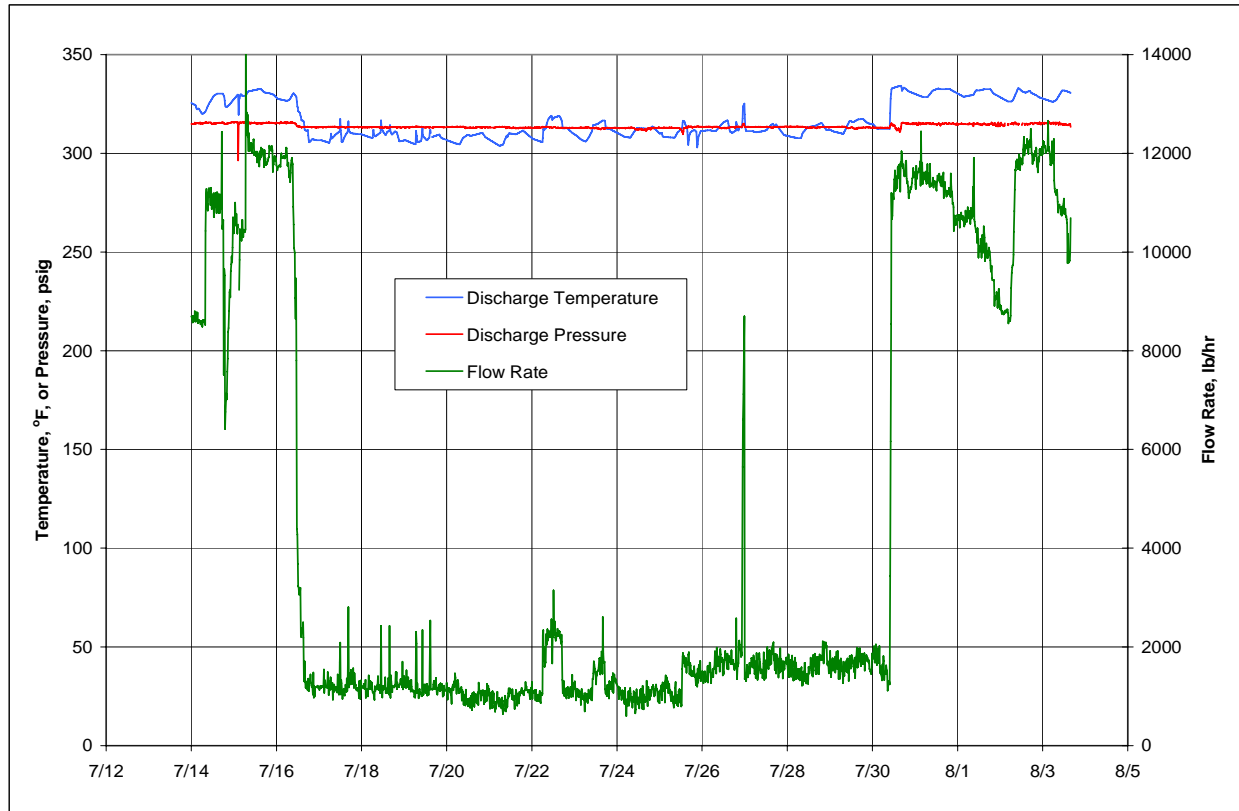


Figure A4-9 Main Air Compressor Operation, 7/14/04 through 8/4/04

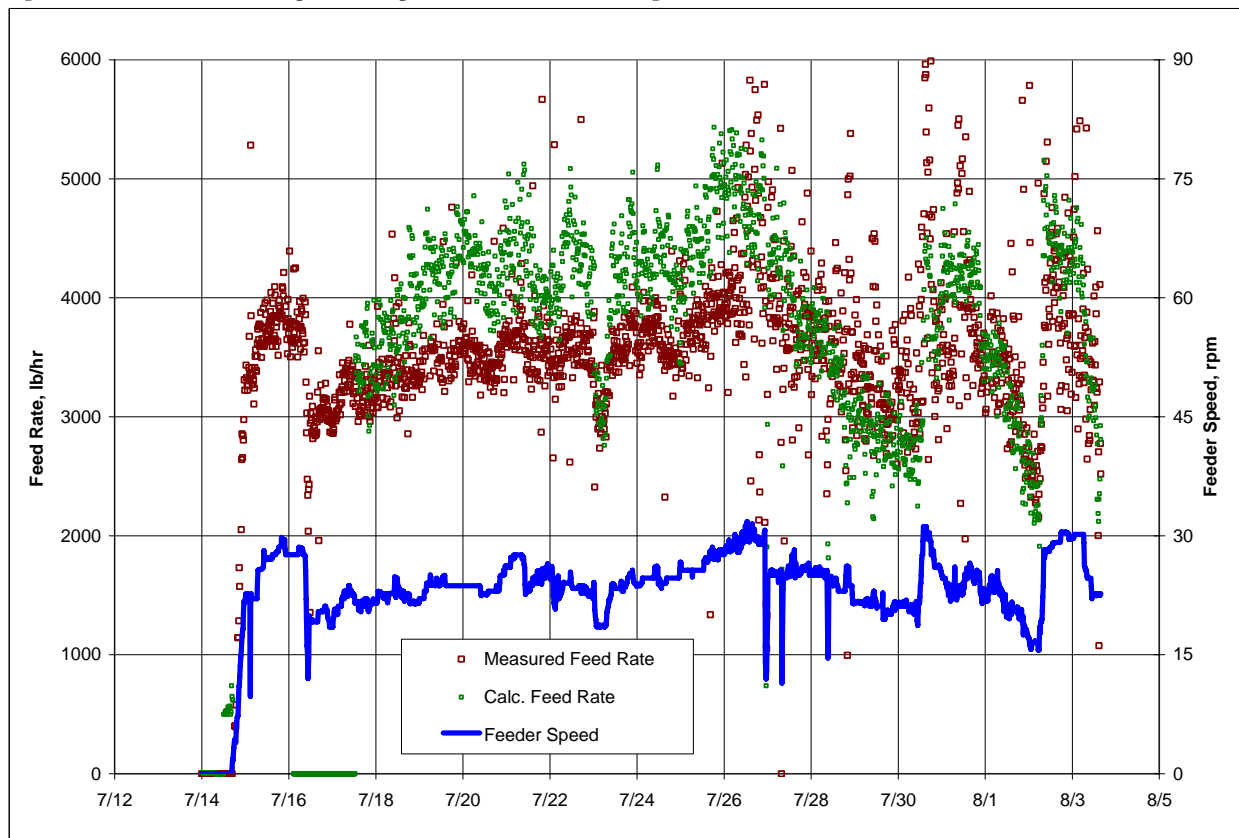


Figure A4-10 Original Coal Feeder Operation, 7/14/04 through 8/4/04

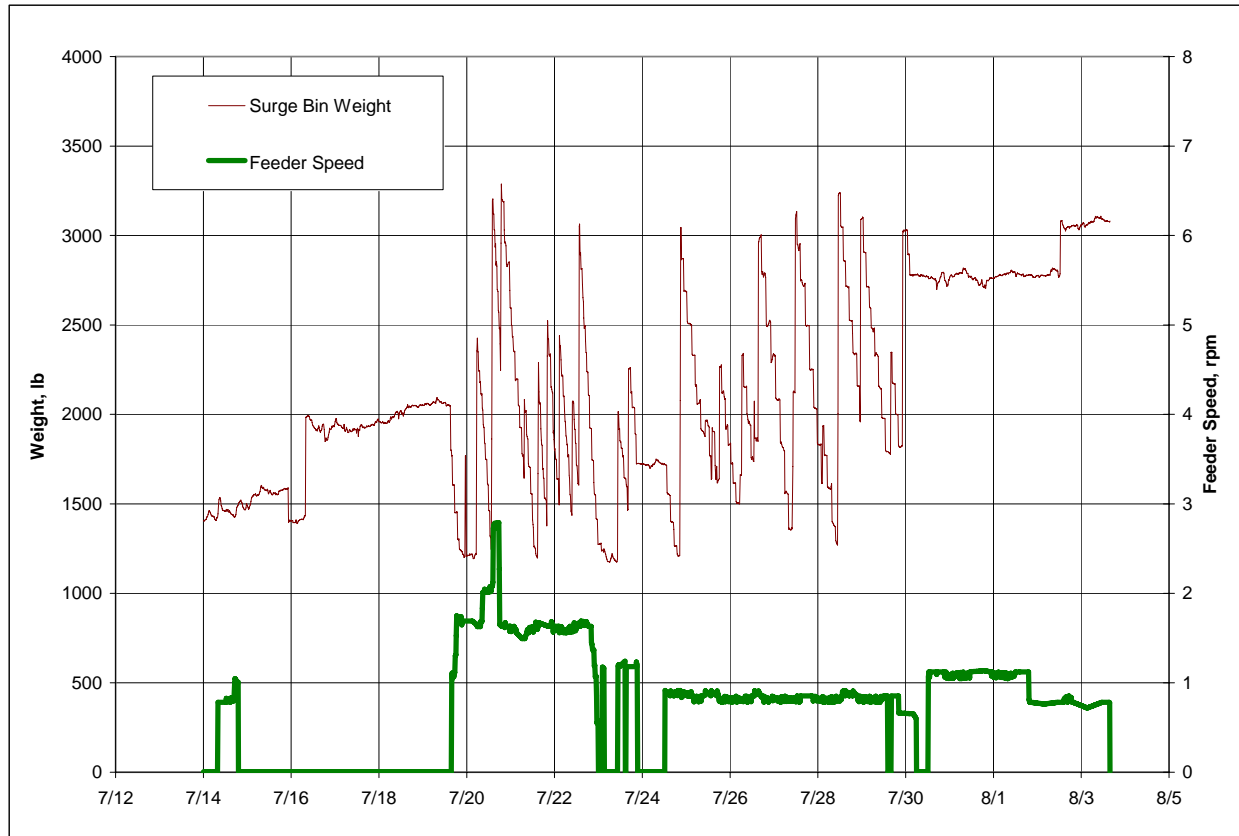


Figure A4-11 Sorbent Feeder Operation, 7/14/04 through 8/4/04

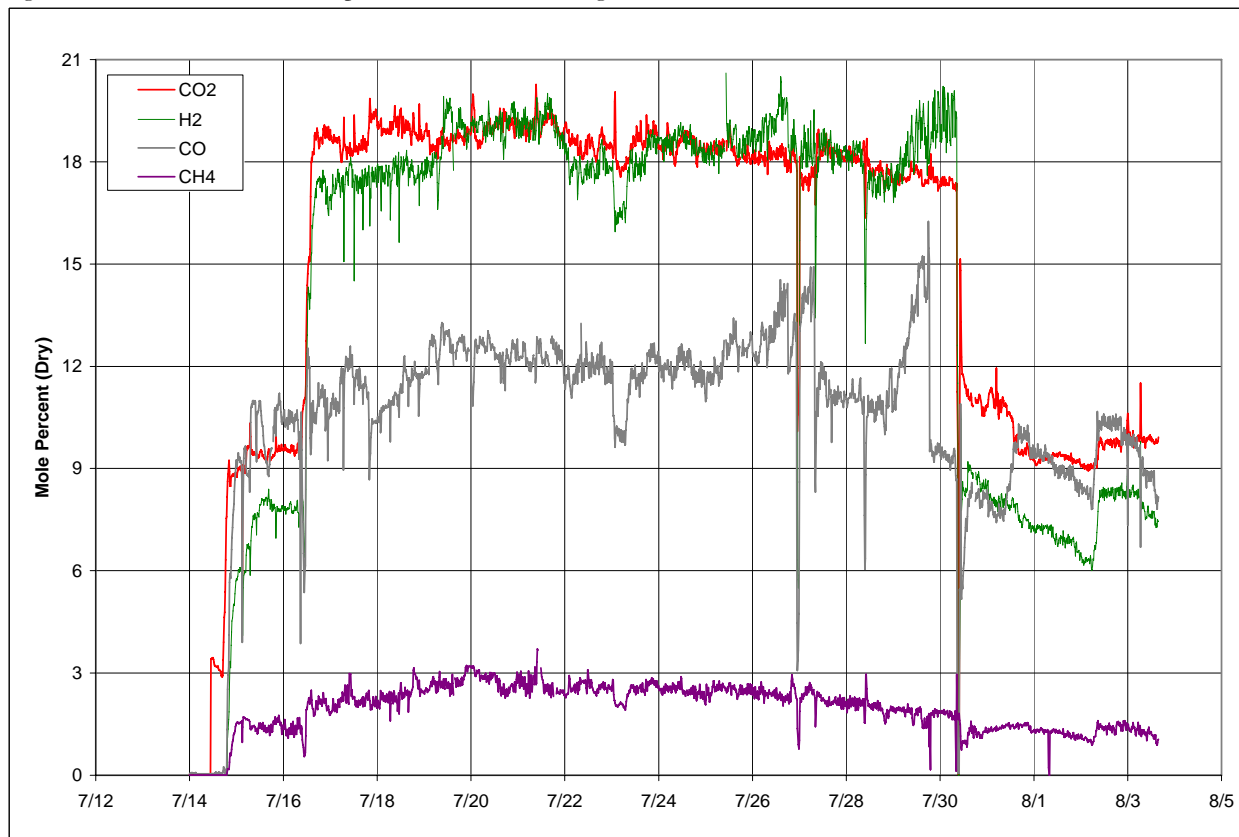


Figure A4-12 Syngas Analyzers, 7/14/04 through 8/4/04

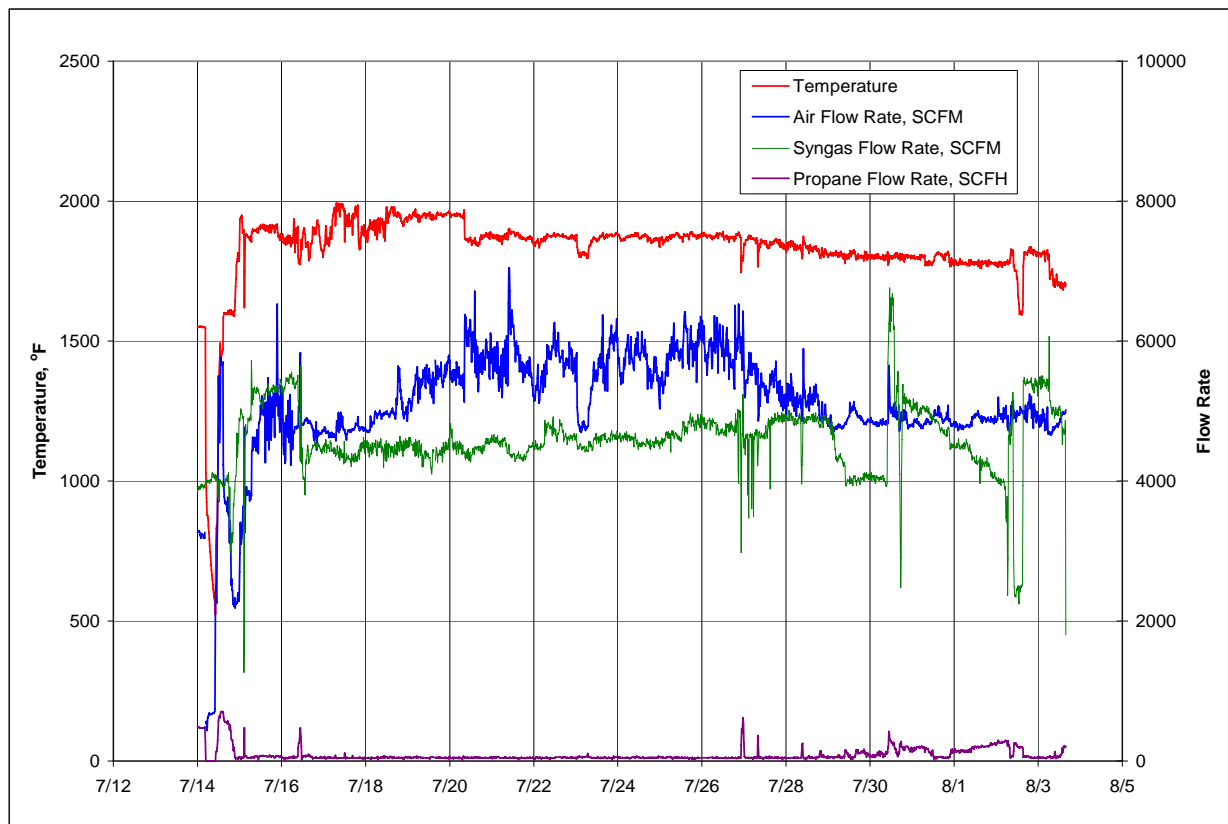


Figure A4-13 Atmospheric Syngas Combustor Operation, 7/14/04 through 8/4/04

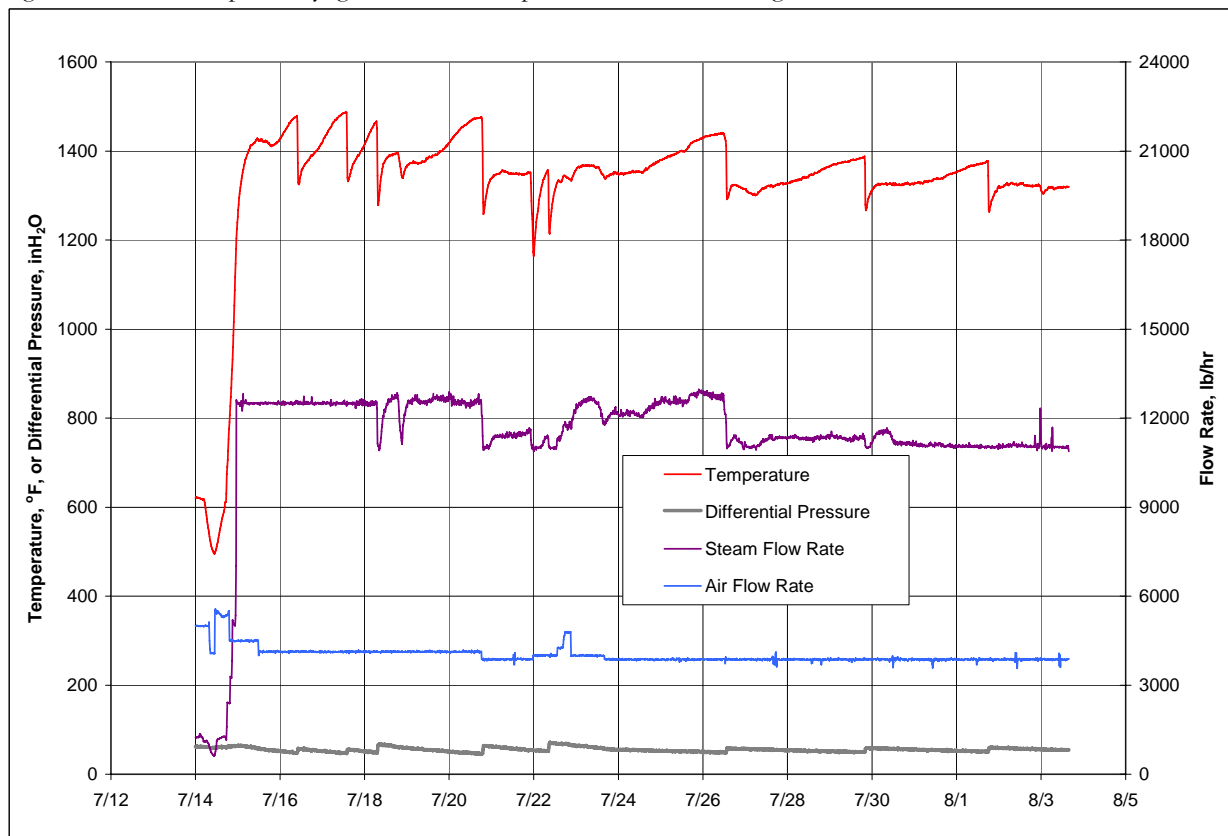


Figure A4-14 Fluidized Bed Combustor Operation, 7/14/04 through 8/4/04

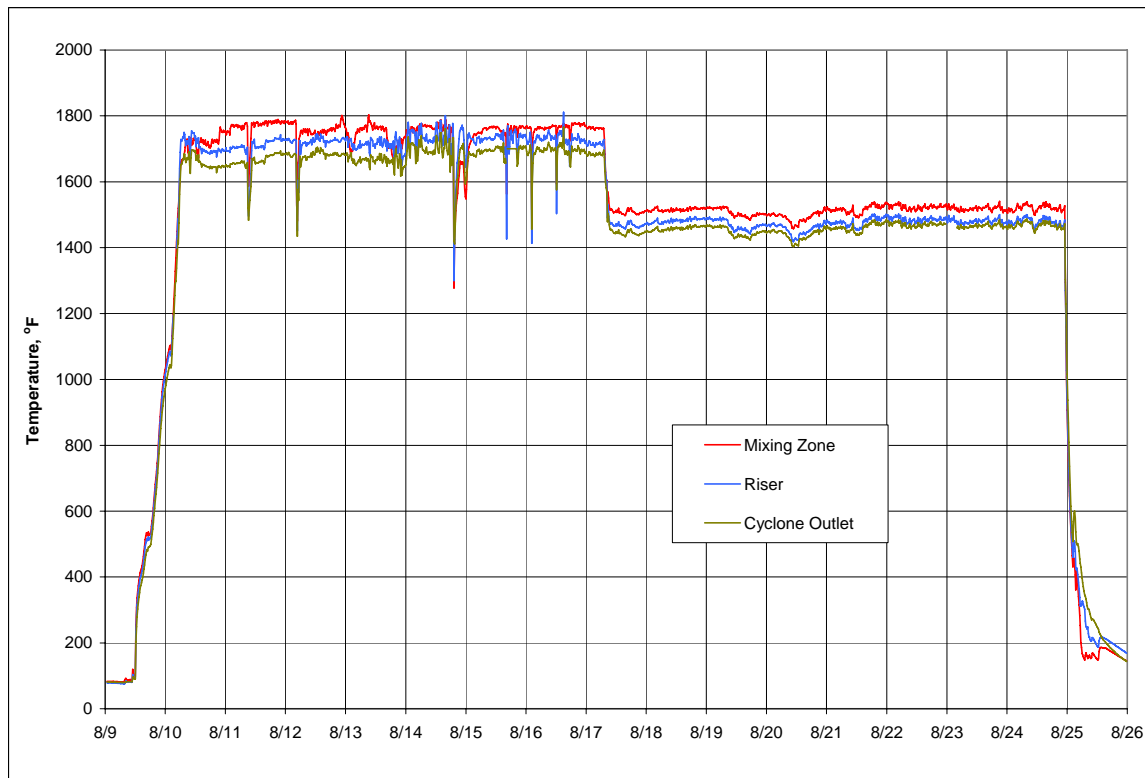


Figure A4-15 Gasifier Mixing Zone, Riser, and Outlet Temperatures, 8/9/04 through 8/26/04

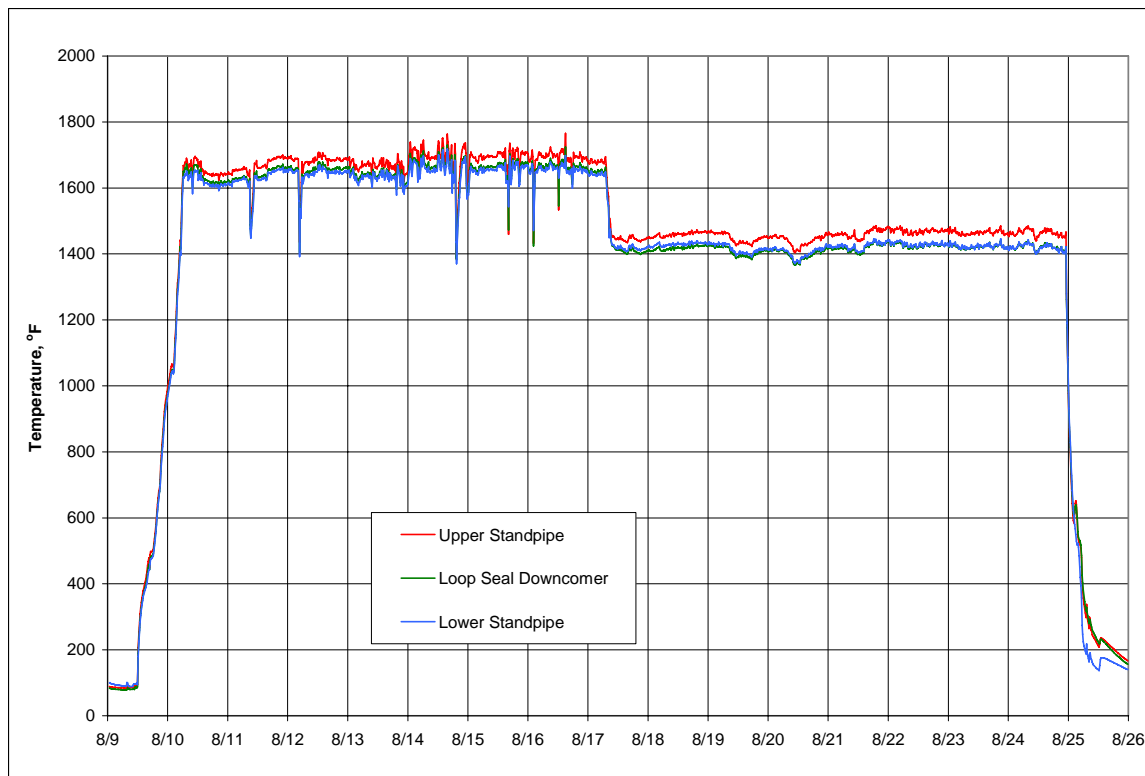


Figure A4-16 Standpipe and Loop Seal Temperatures, 8/9/04 through 8/26/04

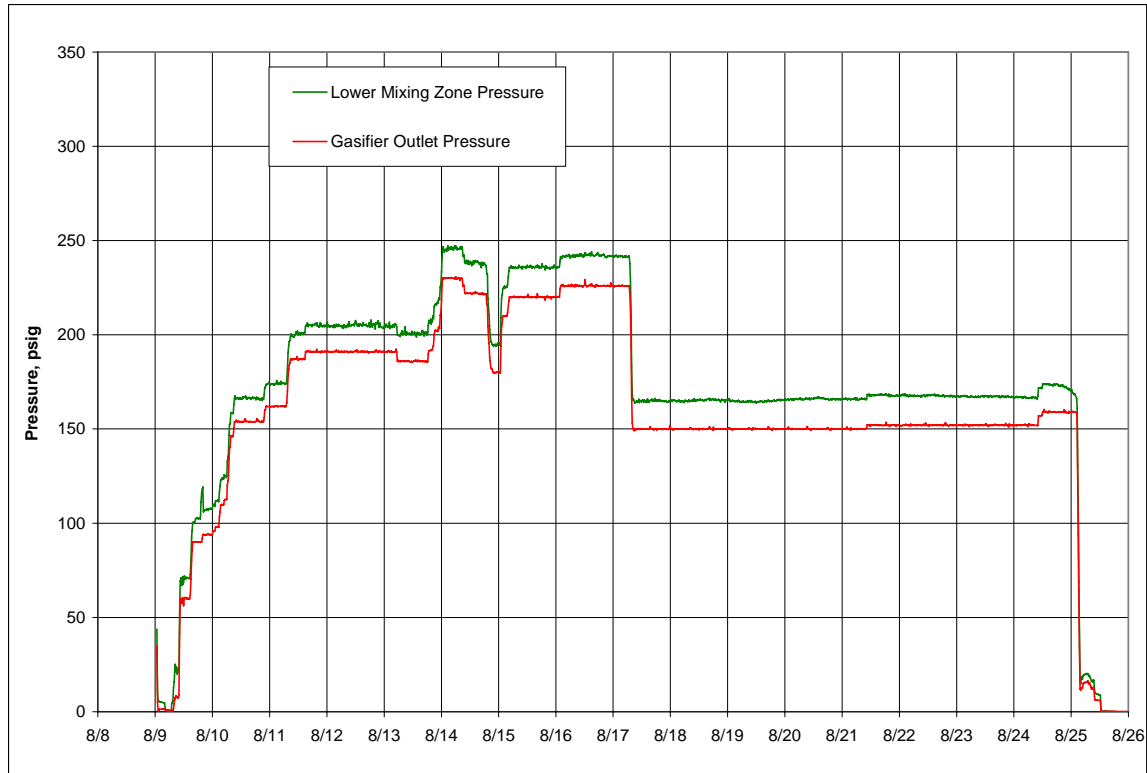


Figure A4-17 Gasifier Pressures, 8/9/04 through 8/26/04

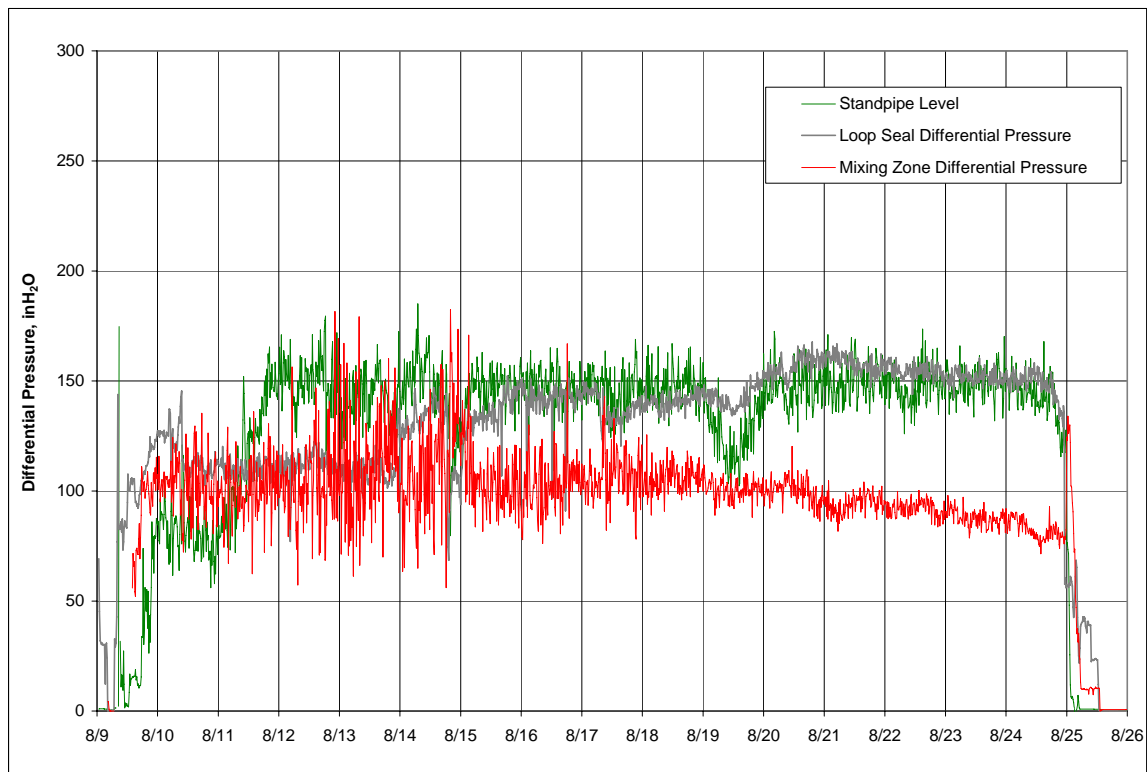


Figure A4-18 Gasifier Differential Pressures, 8/9/04 through 8/26/04

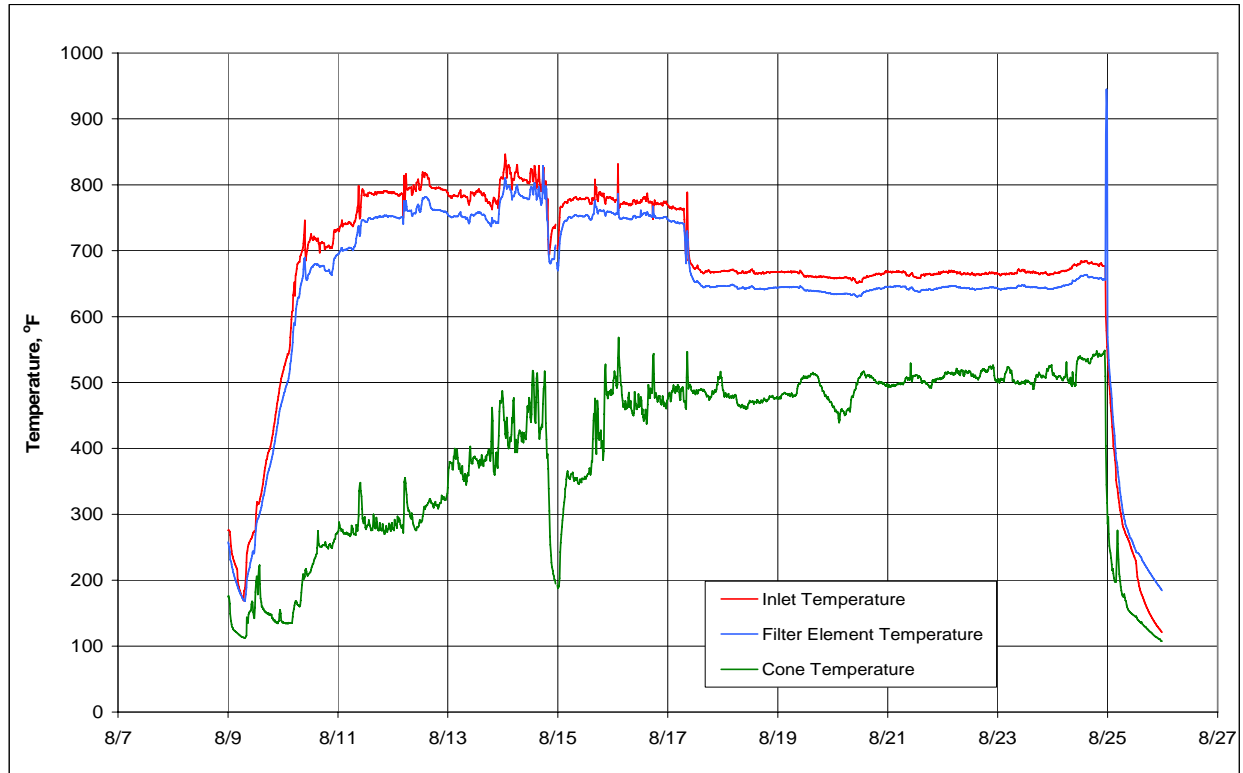


Figure A4-19 PCD Temperatures, 8/9/04 through 8/26/04

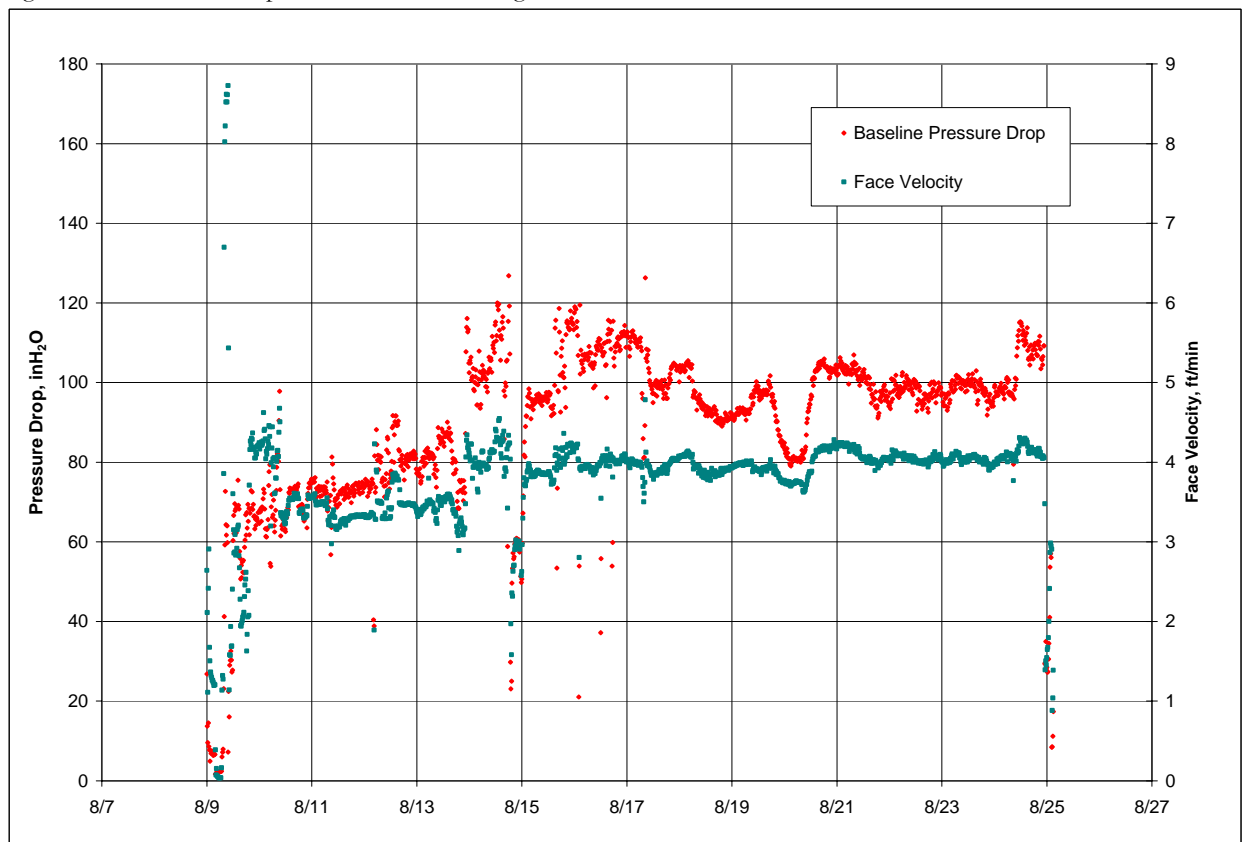


Figure A4-20 PCD Baseline Pressure Drop and Face Velocity, 8/9/04 through 8/26/04

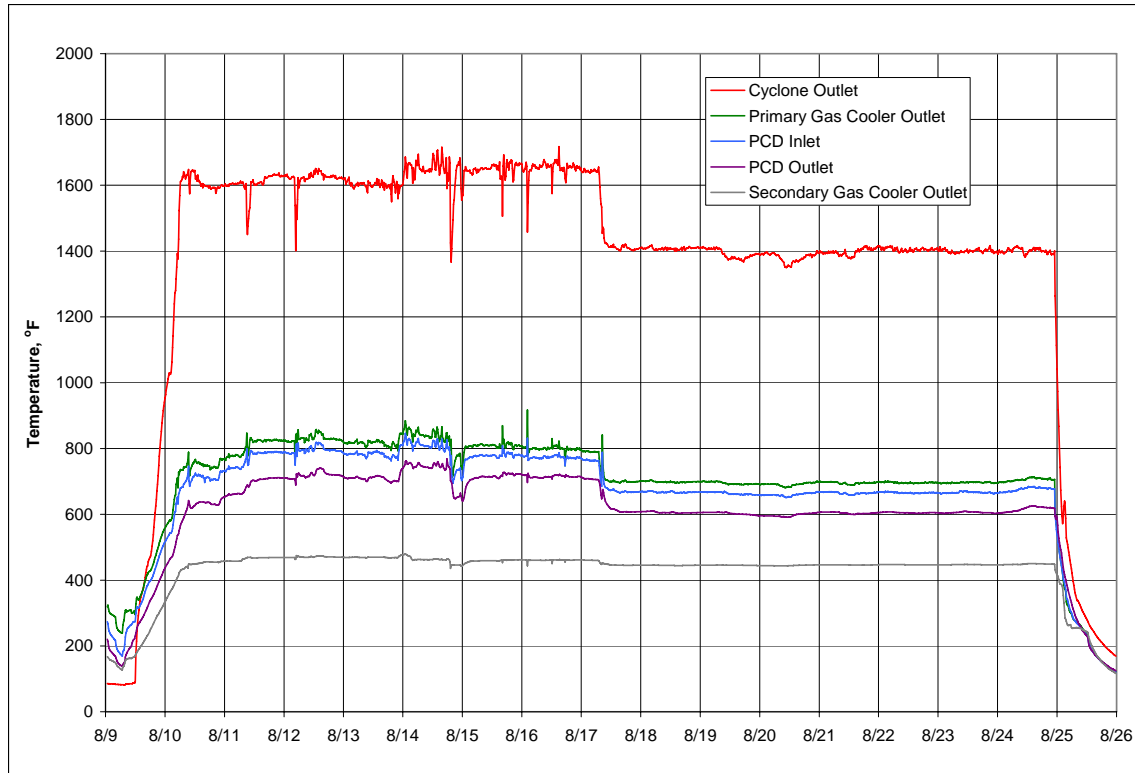


Figure A4-21 System Temperature Profile, 8/9/04 through 8/26/04

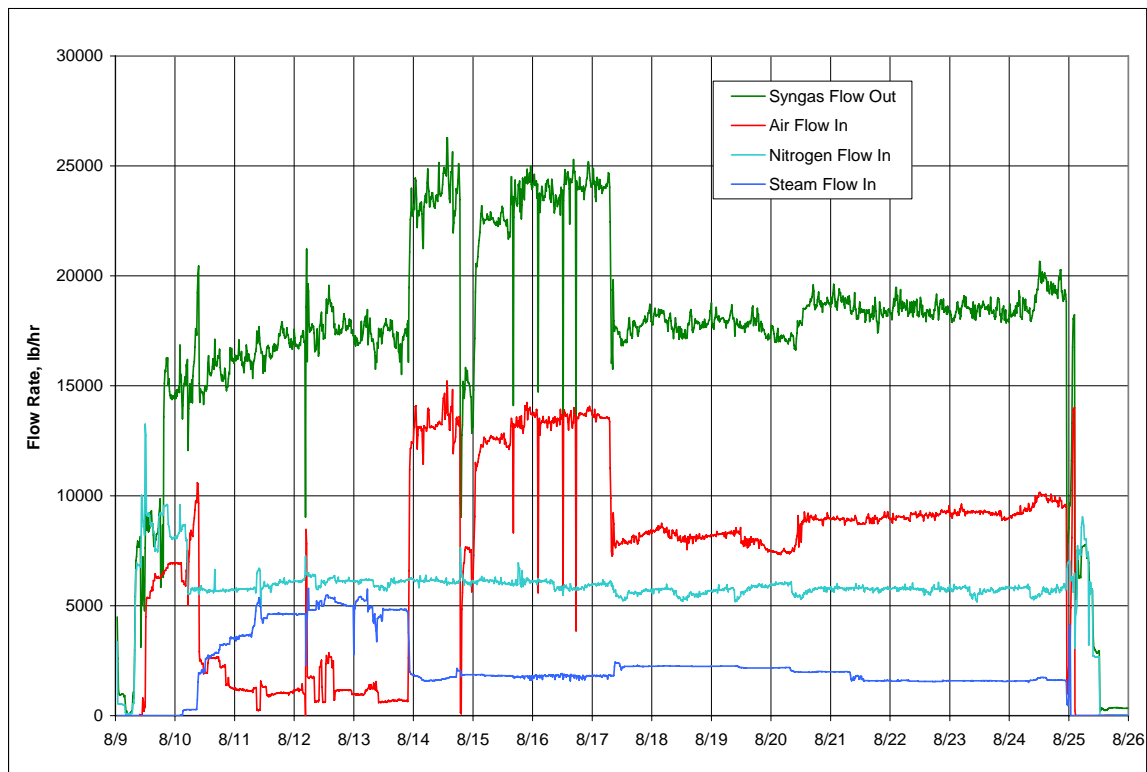


Figure A4-22 System Gas Flows, 8/9/04 through 8/26/04

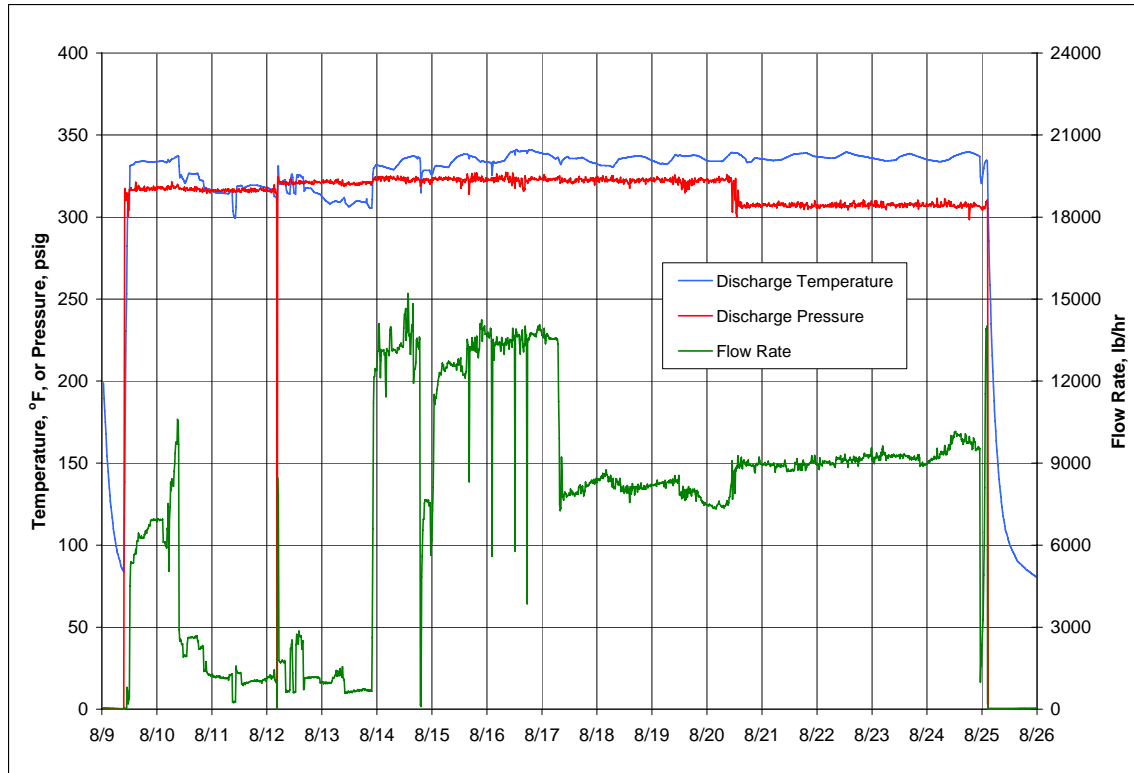


Figure A4-23 Main Air Compressor Operation, 8/9/04 through 8/26/04

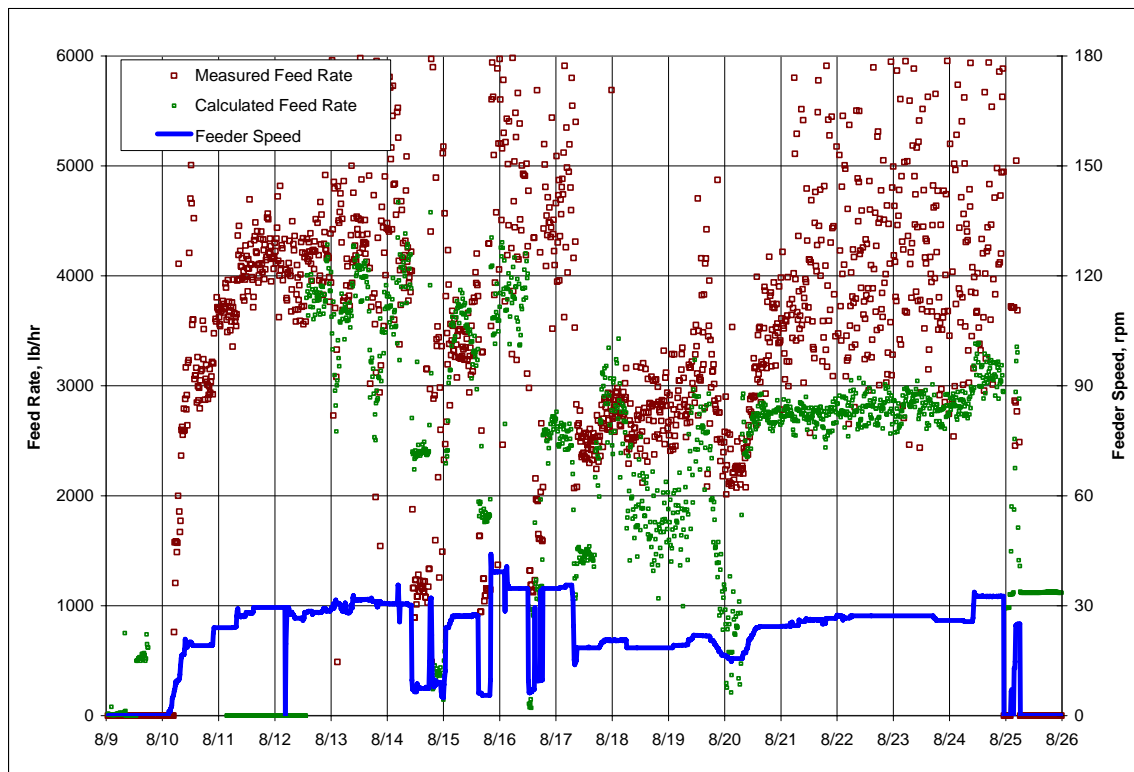


Figure A4-24 Original Coal Feeder Operation, 8/9/04 through 8/26/04

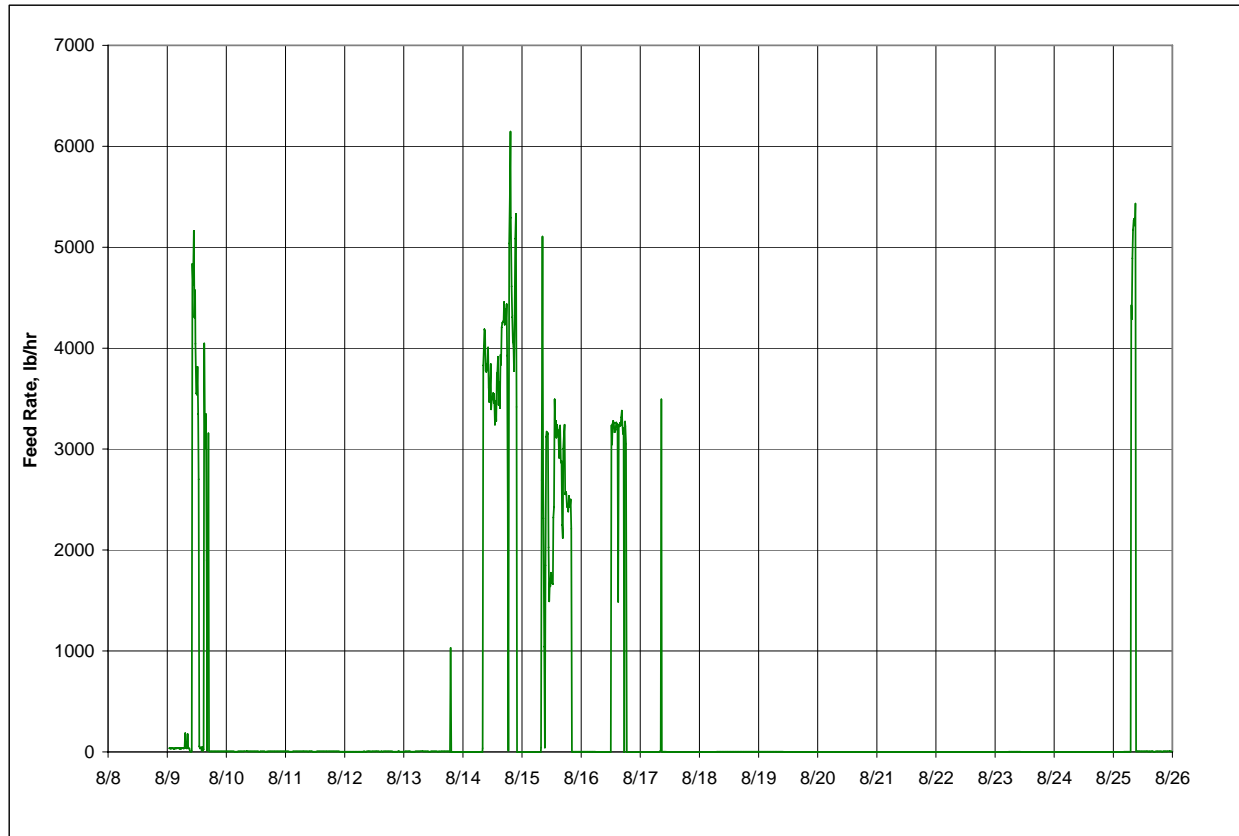


Figure A4-25 Developmental Coal Feeder Operation, 8/9/04 through 8/26/04

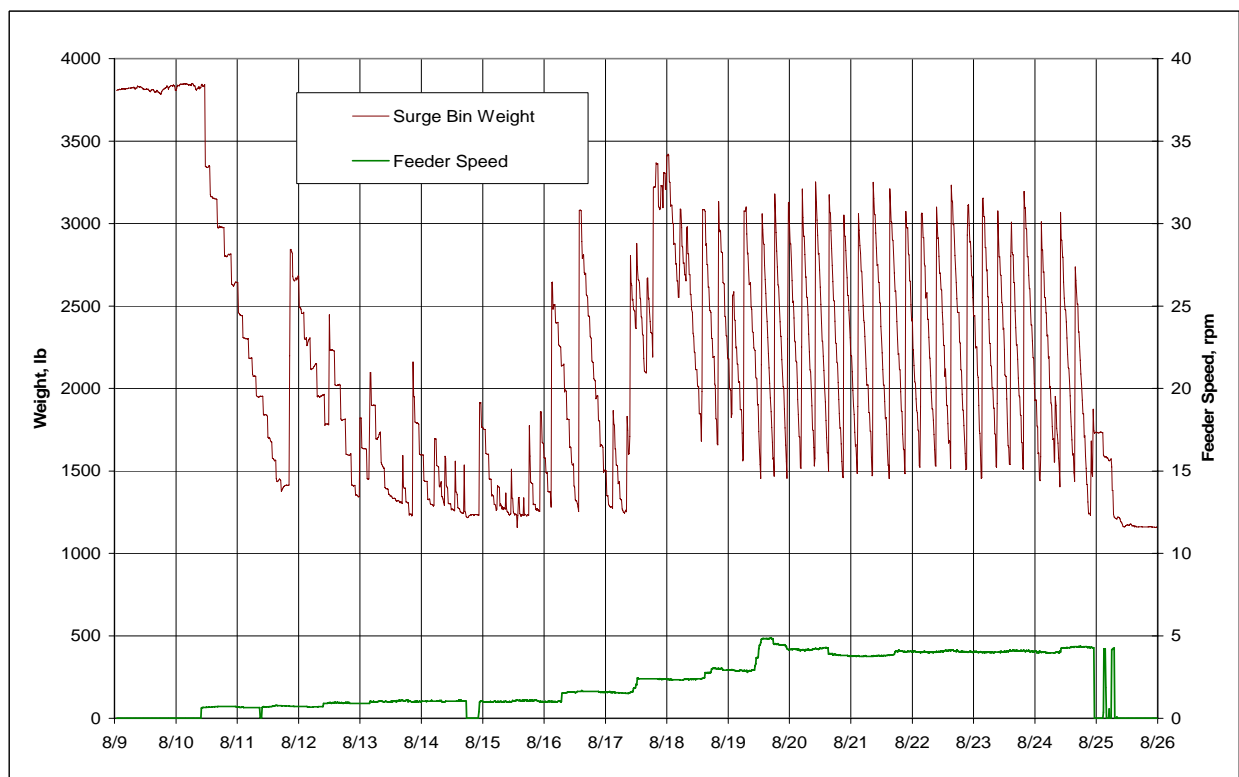


Figure A4-26 Sorbent Feeder Operation, 8/9/04 through 8/26/04

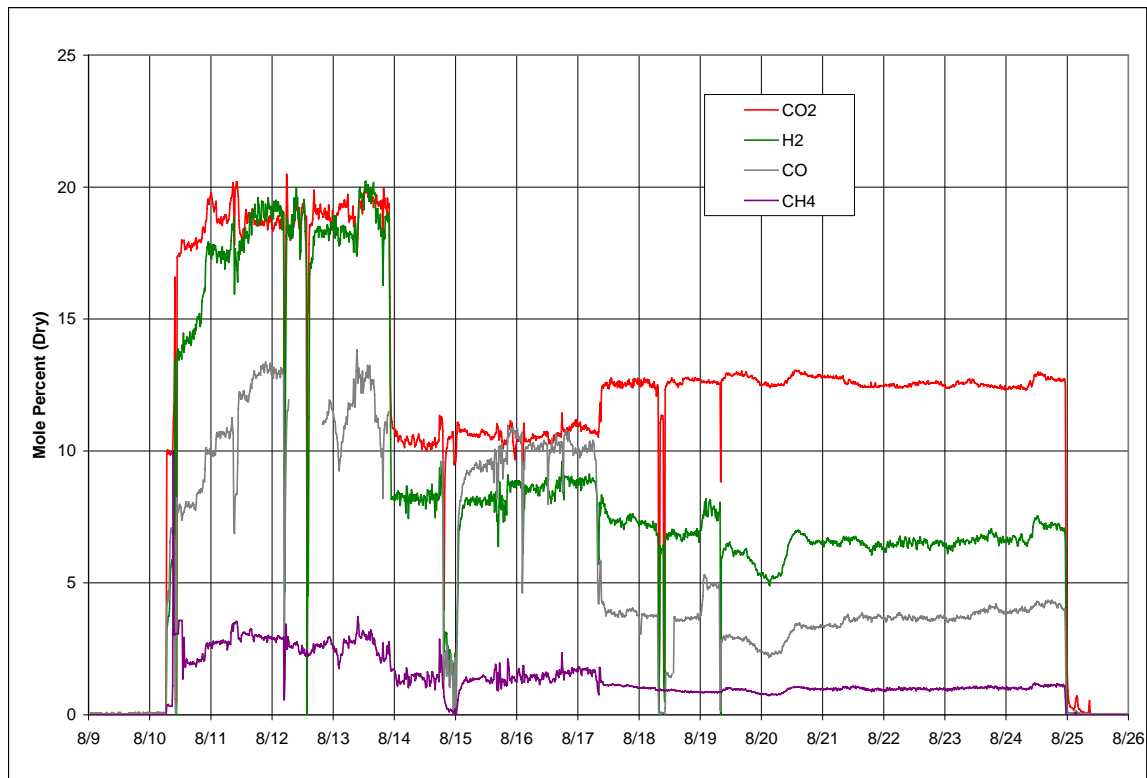


Figure A4-27 Syngas Analyzers, 8/9/04 through 8/26/04

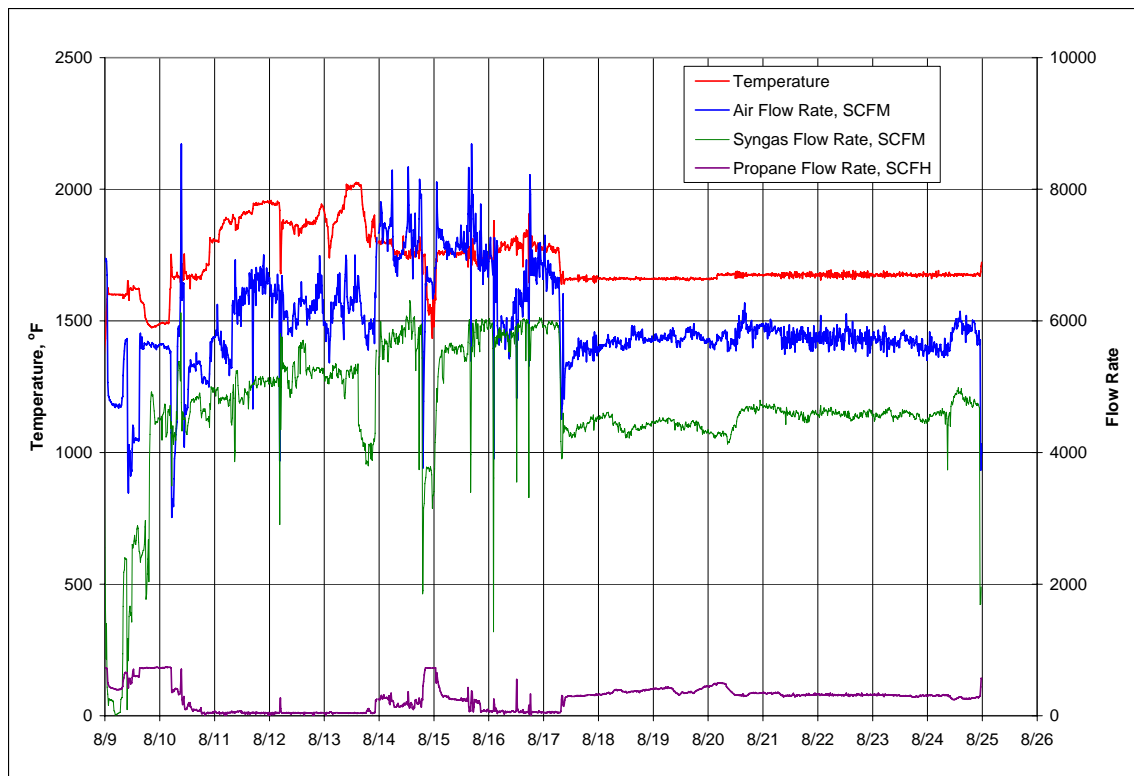


Figure A4-28 Atmospheric Syngas Combustor Operation, 8/9/04 through 8/26/04

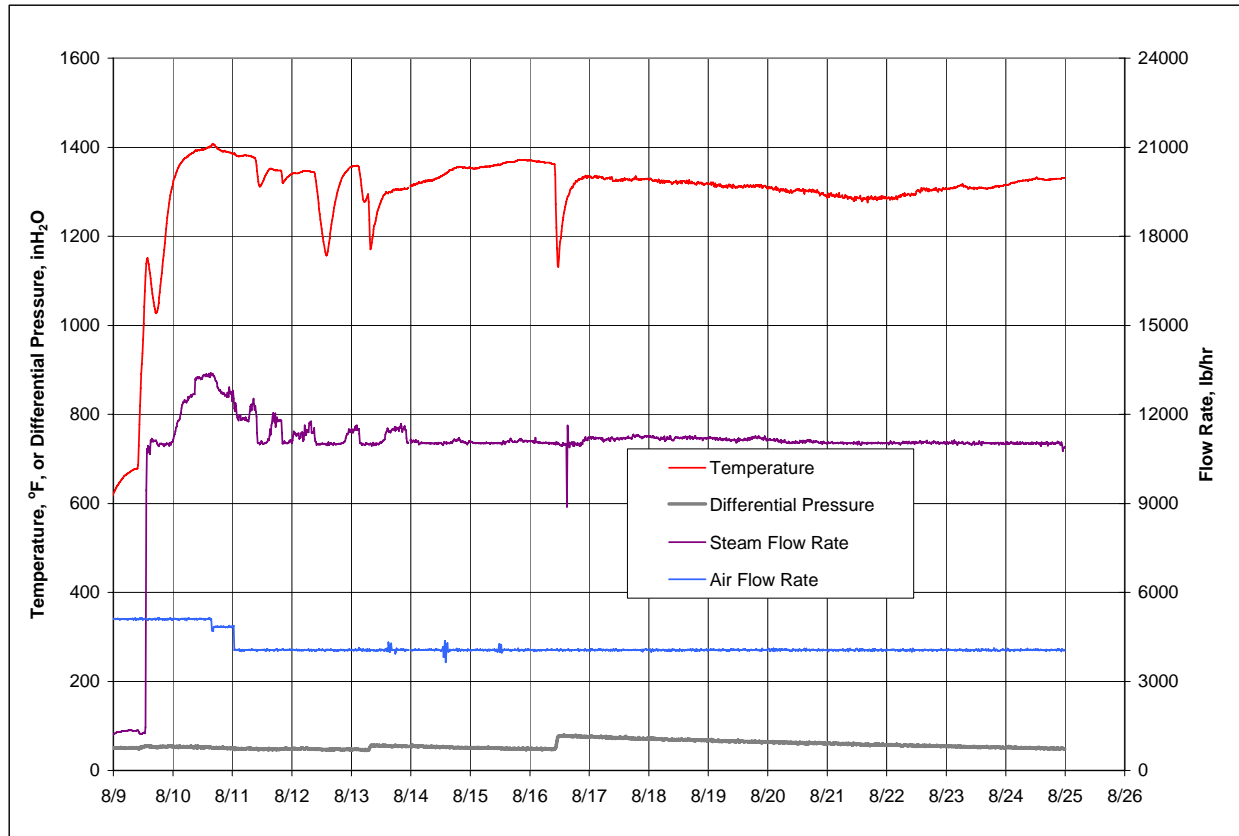


Figure A4-29 Fluidized Bed Combustor Operation, 8/9/04 through 8/26/04

APPENDIX A5 LHV PROJECTION CALCULATIONS

To project a commercial syngas LHV, the following adjustments are made to the raw syngas composition:

1. All non-air nitrogen is removed from the syngas. A commercial plant will have substantially less instrumentation than the PSDF. Because each individual instrument in a commercial plant will require the same purge flow rate as the corresponding instrument at the PSDF, the total instrument purge flow rate will be less. It is assumed that recycled syngas will be used in a commercial plant for aeration. This correction has the effect of increasing all the non-nitrogen syngas compositions and decreasing the nitrogen syngas composition. The recycle syngas flow enters the compressor after the cold gas cleanup system. Since the total amount of nitrogen entering the system is reduced, less coal energy will be required to heat the nitrogen and the coal and air/oxygen feed rates will decrease accordingly. It is assumed that this coal would have been combusted to CO_2 and H_2O . Eliminating this additional coal reduces the syngas CO_2 and H_2O concentrations. The lower projected air rates for air-blown mode also decrease the nitrogen content in the projected syngas, and thus decreases the syngas flow rate. The CO/CO_2 ratio will change due to the reduction in CO_2 . This calculation requires an estimated recycle gas flow rate and an estimated steam aeration rate to determine the heat required to heat the recycle gas to system temperature. The recycle gas flow rate is estimated to be 2.4 percent of the syngas flow rate from the gasifier and is available at 235°F. The aeration steam flow rate is estimated to be 1.45 percent of the syngas flow rate from the gasifier and available at 660°F.
2. Small scale pilot and demonstration units, such as the PSDF, have higher surface area to volume ratios than their scaled up commercial counterparts. Since the heat loss of a commercial plant is difficult to estimate, the projected heat loss is assumed to be zero (adiabatic). The coal, air, and oxygen rates are reduced; the syngas CO_2 , H_2O , and N_2 concentrations are reduced; the CO/CO_2 ratio change. Based on energy balance data, the heat loss for the PSDF Transport gasifier is approximately 5 million Btu/hr.
3. The steam flow rate is adjusted. The steam-to-oxygen ratio will be the same for the PSDF and the commercial Transport Gasifier. Since Steps 1 and 2 reduce the amount of oxygen required, the steam flow rate will decrease correspondingly. The effect of lowering the steam rate will decrease the amount of H_2O in the syngas by the amount the steam rate was reduced. The steam rate and the H_2O content of the syngas are reduced, and hence, the LHV also changes.
4. The water gas shift is recalculated to reflect the gasifier exit temperature. Corrections #1, #2, and #3 change the water gas shift equilibrium constant without affecting the gasifier exit temperature. The commercial plant will operate at the same gasifier exit temperature as the PSDF and hence have the same water gas shift equilibrium constant. The H_2O , CO_2 , CO , and H_2 concentrations are then adjusted based on the water gas shift equilibrium for the temperature of that particular operating period. The LHV could increase if H_2 and CO_2 are converted to H_2O and CO , since the LHV for CO is higher than for H_2 . The LHV will decrease if H_2O

and CO are converted to H₂ and CO₂. The LHV correction is usually small, but the change in composition is important if the syngas is used in a fuel cell or for chemical production where the H₂ concentration is a critical design parameter.

5. The commercial plant will use a cold syngas cleanup train that will drop the syngas temperature to 150°F, before being reheated prior to entering the gas turbine. At these conditions moisture will condense from the syngas and exit via a liquid stream. For the commercial design at 388 psia, the syngas water composition at the gas turbine inlet is 0.96 percent. Thus, the final step reduces the syngas moisture content to this value and adjusts the other contents accordingly.

The result of all of these corrections is the commercially projected LHV. Changes #1 and #2 both increase the oxygen-blown LHV more than for the air-blown LHV because 100 percent of the syngas nitrogen is removed in the oxygen-blown projection, while only about 50 percent of the syngas nitrogen is removed for the air-blown projection.

These calculations are an oversimplification of the gasification process. A more sophisticated model is required to precisely predict the effects of decreasing pure nitrogen and gasifier heat loss. Note that the projected syngas compositions are based on a projected coal rate, projected air rate, projected oxygen rate, projected steam rate, and a projected syngas rate.